Selective Catalytic Oxidation of Hydrogen Sulfide for Simultaneous Coal Gas Desulfurization and Direct Sulfur Production (SCOHS)

Systems Analysis

Letter Report
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Project Manager:
M. D. Rutkowski

Principal Investigators:
T. L. Buchanan
M. G. Klett
R. L. Schoff

PARSONS

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LIST OF ACRONYMS AND ABBREVIATIONS

AGR	acid gas removal
atm	atmosphere
°C	degrees Centigrade
CO	carbon monoxide
COE	cost of electricity
COS	carbonyl sulfide
°F	degrees Fahrenheit
ft	foot
GT	gas turbine
h	hour
H_2S	hydrogen sulfide
HCN	hydrogen cyanide
HGD	hot gas desulfurization
HHV	high heating value
HRSG	heat recovery steam generator
ID	inside diameter
IGCC	integrated gasification combined cycle
kW	kilowatt
lb	pound
LP	low pressure
MDEA	monodiethanolamine
MMBtu	million British thermal units
MWe	megawatt electric
MWh	megawatt hour
NETL	National Energy Technology Laboratory
NH_3	ammonia
NOx	oxides of nitrogen
ppmv	parts per million by volume
psia	pound per square inch absolute
scf	standard cubic foot
SCOHS	selective catalytic oxidation of hydrogen sulfide
SCR	selective catalytic reduction
sec	second
SNCR	selective non-catalytic reduction
SO_2	sulfur dioxide
STP	standard pressure and temperature
TECO	Tampa Electric Company
tpd	tons per day

EXECUTIVE SUMMARY

INTRODUCTION

The selective catalytic oxidation of hydrogen sulfide (SCOHS) process may have the potential of removing hydrogen sulfide (H_2S) from syngas and producing elemental sulfur in one step through development of an intrinsically simpler cleanup system than both the NETL-sponsored hot gas desulfurization (HGD) and the current state-of-the-art amine and methanol technology. This is due to several salient points such as the reduction in the overall number of process steps and, correspondingly, a reduction in process complexity. Both hot gas and conventional amines require separate sulfur removal and sulfur production facilities. The SCOHS process takes advantage of the selective oxidation of H_2S to sulfur with oxygen injected directly into the syngas stream, and can condense the sulfur from the syngas in one process. The exact chemistry of this process may be represented by the following reaction:

$$H_2S + \tfrac{1}{2}O_2 \rightarrow \tfrac{1}{n}S_n + H_2O \qquad (1)$$

where n = 2, 6, or 8, depending on the temperature of the reaction.

The objective of this task is to perform a systems analysis of the production of clean syngas, comparing the relative performance and economics of conceptual plant concepts at low and medium temperatures for sulfur recovery, with the medium temperature being the SCOHS process. Parsons has completed a reference Texaco coal gasification plant design, based on the Tampa Electric Company (TECO) IGCC Demonstration Plant,[1] which will be utilized as the starting point for the analysis.

The following equipment will be removed from the TECO plant:

- Sour gas coolers and knockout drum.
- MDEA unit and clean gas reheat heat exchanger.
- Claus plant.
- Tail gas treatment unit and tail gas incinerator.

Figure ES-1 is the block flow diagram of the IGCC plant with the SCOHS process replacing the conventional sulfur removal processes.

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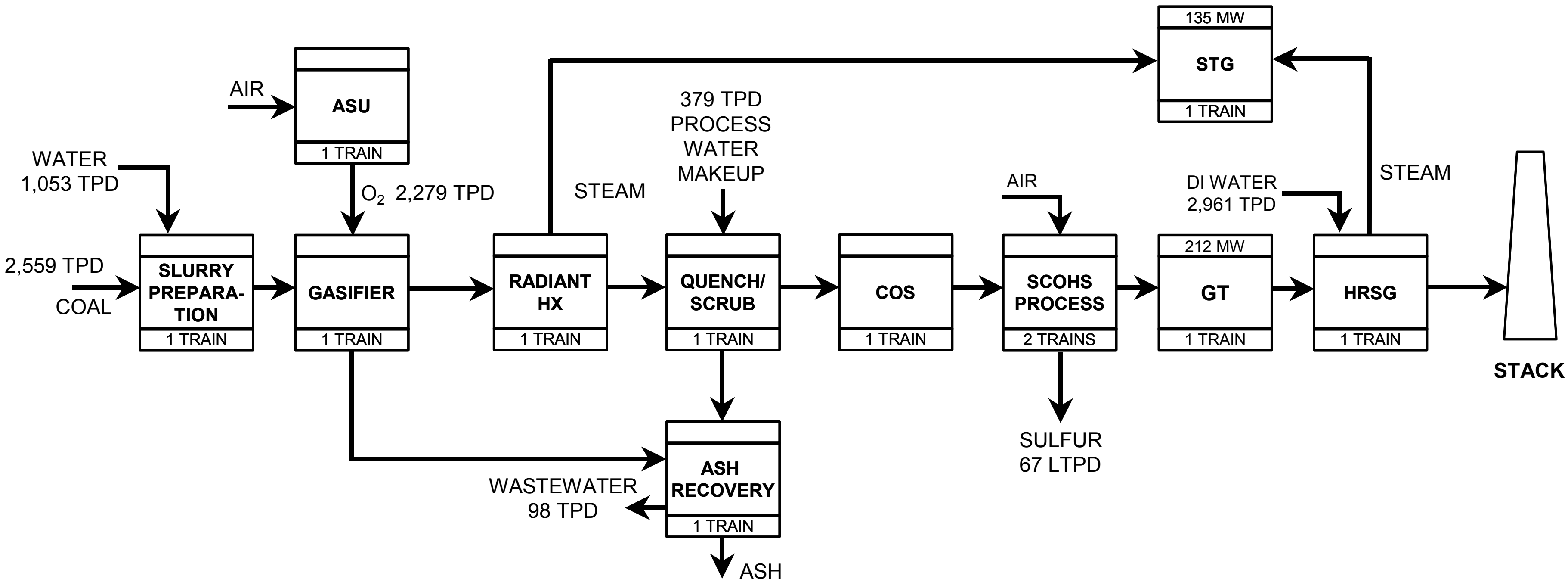
[1] "Clean Coal Reference Plants: Integrated Gasification Combined Cycle, Texaco," prepared for the United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AM26-99T40465, Draft Report, January 2001.

Figure ES-1
Block Flow Diagram – Texaco Radiant Cooler IGCC Plant with Fixed-Bed SCOHS Process



CASE 1 – SCOHS FIXED BED

For the fixed-bed cases, desulfurization of the fuel gas is accomplished in a fixed bed of granular activated carbon catalyst. The activated carbon catalyst is in the form of 1/8 to 3/16-inch-diameter granules. Based on experimental data from Todd H. Gardner at the U.S. Department of Energy, National Energy Technology Laboratory, it was assumed that all SCOHS reactors have only 20 ppmv COS in exit, independent of inlet H_2S. Therefore, with 8,441 ppmv H_2S inlet, 20 ppmv COS in outlet results in 99.76 percent recovery.

The catalyst loading and gas space velocity determine the initial time to breakthrough for a fuel gas with a specified sulfur concentration. For this study, it was assumed that the catalyst loading was 50 percent at a space velocity of 1,000 hr^{-1} (STP). This results in a 24-hour absorption cycle and, assuming a catalyst life of 100 cycles, a catalyst replacement time of 6.6 months.

Since the sulfur reactions, which occur during absorption, are exothermic, bed temperature is controlled to a maximum of 310°F (the turbine inlet temperature) by maintaining the syngas temperature at 275°F.

For the SCOHS baseline design, regeneration of the catalyst bed is required when the catalyst reaches 50 percent of its weight with sulfur. The method selected for catalyst regeneration includes heating the bed and extracting sulfur vapors with hot circulating nitrogen. In order to regenerate the catalyst, it must be heated to a temperature at which sulfur vapor pressure increases and the vapors can be swept away. Figure ES-2 shows the schematic block flow diagram for the regeneration step. By appropriate valving, a continuous nitrogen loop sends nitrogen through the bed, through a regenerative heat exchanger, and through a sulfur condenser. Following the sulfur condenser, liquid sulfur can be removed from the gas/liquid separator. The nitrogen is then recirculated with a boost compressor.

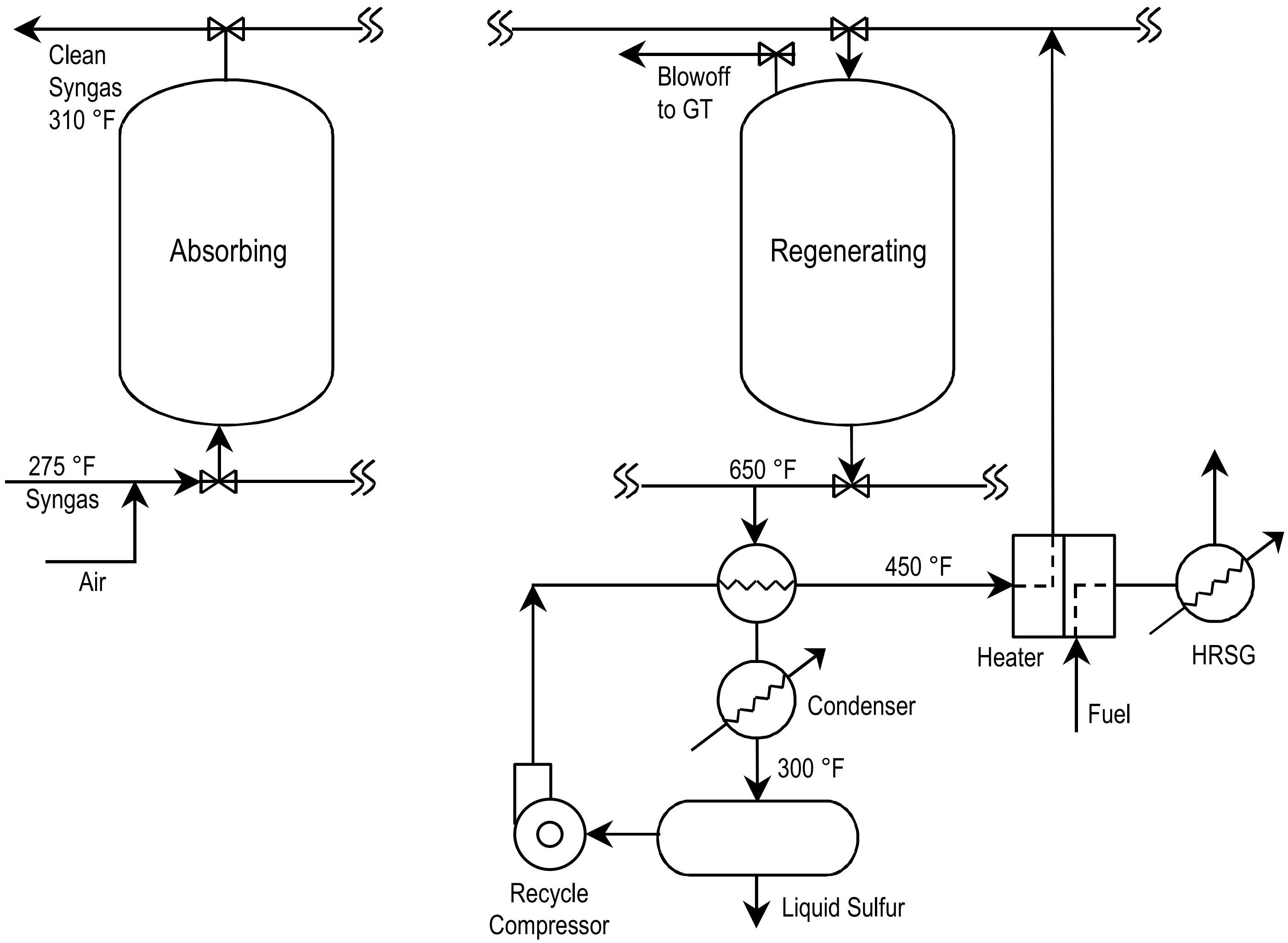
It was determined that the best temperature for regeneration is 650°F, at which point the vapor pressure of sulfur is 2.8 psia. This is a reasonable temperature for control and material selection, and will permit full regeneration in the available time frame.

Following catalyst regeneration, the bed is allowed to cool back to 275°F by continuing to circulate nitrogen and extract heat using the sulfur condenser. The sulfur condenser has a capacity of 3.12 MMBtu/hour, which results in bed cooling to 275°F in 5 hours.

The cool catalyst bed, free of sulfur, is maintained at 275°F for approximately 1 hour. It is then switched to the active syngas stream for the next absorbing cycle, and the parallel catalyst vessel is taken off stream to be regenerated.

The fixed-bed SCOHS retrofit of the Tampa Electric IGCC Demonstration Project results in a plant producing a net output of 301 MWe at a net efficiency of 38.7 percent on an HHV basis. Performance is based on the properties of Pittsburgh No. 8 coal.

Figure ES-2
Flow Diagram – Fixed-Bed SCOHS Retrofit



The overall capital cost difference between the plants is small, notwithstanding the significant cost changes in the sulfur removal and sulfur recovery processes. Table ES-1 shows the factored adjustment of capital from the Texaco-based IGCC Reference Plant to the plant with fixed-bed SCOHS retrofit. The initial estimate for the cost of the SCOHS equipment is nominally $10 million. Although the absolute plant total capital requirement changed by only 2 percent, the SCOHS capital cost per kW is lowered by 6 percent because of higher plant efficiency and higher power production.

Table ES-1
Capital Cost Changes with Fixed-Bed SCOHS Retrofit

	Texaco IGCC Plant Size 287 MWe		Fixed-Bed SCOHS Plant Size 301 MWe	
	1,000$	**$/kW**	**1,000$**	**$/kW**
Total Plant Costs	$342,572	$1,195	$336,474	$1,118
Others	$37,353	$130	$36,353	$121
Total Capital Requirements	$379,925	$1,325	$372,827	$1,239

The cost of electricity (COE) for the fixed-bed SCOHS retrofit was determined by adjusting the COE from the Texaco-based IGCC Reference Plant. Table ES-2 shows the changes in the components making up the COE. The COE is based on a fuel cost of $1.25 per MMBtu and an annual plant capacity factor of 80 percent. The change in COE with the fixed-bed SCOHS retrofit amounts to a reduction of 6.4 percent.

Table ES-2
Change in Cost of Electricity with Fixed-Bed SCOHS Retrofit

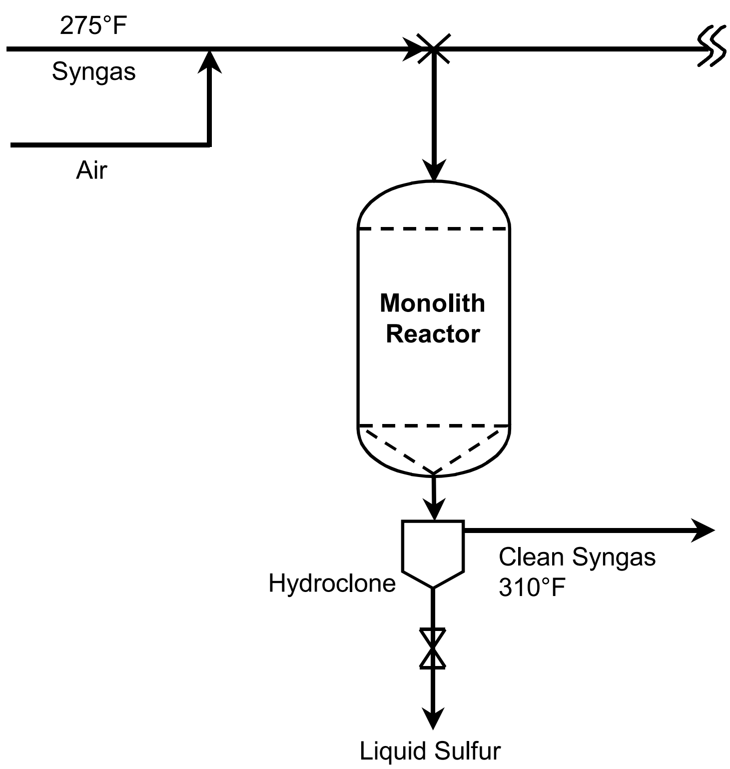
	Texaco IGCC Plant Size 287 MWe		Fixed Bed SCOHS Plant Size 301 MWe	
	1,000$/y	**$/kW-y**	**1,000$/y**	**$/kW-y**
Capital Charge	$52,432	$182.90	$51,450	$171.00
O&M	$12,126	$42.30	$11,626	$38.64
Consumables	$2,332	$8.13	$1,832	$6.09
Sulfur Credit @ $47/LT	($898)	($3.13)	($915)	($3.04)
Fuel @ $1.25/ MMBtu	$22,634	$78.95	$23,081	$76.71
Total	**$88,626**	**$309.16**	**$87,074**	**$287.56**
COE @ 80% CF $/MWh	**44.11**		**41.30**	

CASE 2 – SCOHS MONOLITHIC CATALYST BED

The monolith catalyst version of the SCOHS process utilizes a carbon-fiber-based porous carbon monolithic catalyst (similar to that developed at the Oak Ridge National Laboratory)[2] to continuously catalyze the SCOHS reaction and produce liquid sulfur without a regeneration step.

Figure ES-3 is a process flow diagram for the SCOHS monolith reactor.

Figure ES-3
Flow Diagram – Monolith SCOHS Concept



The SCOHS reaction still requires two reactor trains, but since offline regeneration is not necessary, only two vessels instead of four are required. Because of the low bulk density of the monolith composite, only about half as much catalyst by weight is required. Compressed air at 400 percent stoichiometric of the H_2S-to-sulfur oxidation is mixed with the fuel gas leaving the COS reactor at 402 psia and 275°F.

[2] Burchell, T. D. et al., "A Novel Process and Material for the Separation of Carbon Dioxide and Hydrogen Sulfide Gas Mixtures," Carbon Vol. 35(9), pp. 1279-1294 (1997).

The gas enters the reactor, and the H_2S and oxygen react to form liquid elemental sulfur. The sulfur is carried through the composite monolith by gas flow and gravity to a sump at the bottom of the reactor vessel. The exothermic sulfur reaction causes the gas temperature to increase to 315°F. A high-efficiency hydrocyclone at the bottom of the vessel separates the liquid sulfur from the gas. Because of the monolith and hydrocyclone pressure drop, the fuel gas going to the gas turbine is reduced to 350 psia. Sulfur is continuously drained from the sump at the bottom of the hydrocyclone and is stored for shipment.

The monolith SCOHS retrofit of the Tampa Electric IGCC Demonstration Project results in a plant producing a net output of 303 MWe at a net efficiency of 39.2 percent on an HHV basis. Performance is based on the properties of Pittsburgh No. 8 coal.

The overall capital cost difference between the plants is small, notwithstanding the significant cost changes in the sulfur removal and sulfur recovery processes. Table ES-3 shows the factored adjustment of capital from the Texaco-based IGCC Reference Plant to the plant with monolith SCOHS retrofit. The initial estimate for the cost of the SCOHS equipment is nominally $5 million. Although the absolute plant total capital requirement changed by less than 4 percent, the monolith SCOHS capital cost per kW is lowered by nearly 9 percent because of higher plant efficiency and higher power production.

Table ES-3
Capital Cost Changes with Monolith SCOHS Retrofit

	Texaco IGCC Plant Size 287 MWe		Monolith SCOHS Plant Size 303 MWe	
	1,000$	**$/kW**	**1,000$**	**$/kW**
Total Plant Costs	$342,572	$1,195	$330,960	$1,093
Others	$37,353	$130	$36,353	$120
Total Capital Requirements	$379,925	$1,325	$367,313	$1,213

The cost of electricity (COE) for the monolith SCOHS retrofit was determined by adjusting the COE from the Texaco-based IGCC Reference Plant. Table ES-4 shows the changes in the components making up the COE. The COE is based on a fuel cost of $1.25 per MMBtu and an annual plant capacity factor of 80 percent. The change in COE with the monolith SCOHS retrofit amounts to a reduction of 8 percent.

Table ES-4
Change in Cost of Electricity with Monolith SCOHS Retrofit

	Texaco IGCC Plant Size 287 MWe		Monolith SCOHS Plant Size 303 MWe	
	1,000$/y	**$/kW-y**	**1,000$/y**	**$/kW-y**
Capital Charge	$52,432	$182.90	$50,689	$167.40
O&M	$12,126	$42.30	$11,626	$38.40
Consumables	$2,332	$8.13	$1,832	$6.05
Sulfur Credit @ $47/LT	($898)	($3.13)	($909)	($3.00)
Fuel @ $1.25/ MMBtu	$22,634	$78.95	$22,932	$75.73
Total	**$88,626**	**$309.16**	**$86,171**	**$284.58**
COE @ 80% CF $/MWh	**44.11**		**40.61**	

CONCLUSIONS

Conceptual design of an IGCC plant retrofitted with both the fixed-bed and monolith SCOHS concepts indicates that favorable performance and economics can be achieved. The fixed-bed SCOHS results in a COE reduction of 6.4 percent, while the monolith SCOHS results in a COE reduction of 8 percent.

1. INTRODUCTION

The advantages of removing hydrogen sulfide (H_2S) from syngas and producing elemental sulfur in one step have been discussed for years. The SCOHS process may have the potential to accomplish this goal through development of an intrinsically simpler cleanup system than both the NETL-sponsored hot gas desulfurization (HGD) and the current state-of-the-art amine and methanol technology. This is due to several salient points such as the reduction in the overall number of process steps and, correspondingly, a reduction in process complexity. Another key to economic or efficiency benefit may be the transition from the reduced form of sulfur, H_2S, to the elemental form, S_8, without going through the energy-dependent oxidation up to sulfur dioxide (SO_2) and reduction back to elemental sulfur.

Both hot gas and conventional amines require separate sulfur removal and sulfur production facilities. The SCOHS process takes advantage of the selective oxidation of H_2S to sulfur with oxygen injected directly into the syngas stream, and can condense the sulfur from the syngas in one process. The exact chemistry of this process may be represented by the following reaction:

$$H_2S + \tfrac{1}{2}O_2 \rightarrow \tfrac{1}{n}S_n + H_2O \tag{1}$$

where n = 2, 6, or 8, depending on the temperature of the reaction. At relatively low temperatures, the sulfur product may be solid sulfur, S_8, and at higher temperatures sulfurous gases, S_2 and S_6, are produced. Separation of the sulfur product may be achieved by quenching the coal-derived synthesis gas stream and recovering the sulfur as a pure condensed phase. A number of other undesirable side reactions may also occur. If reaction (1) occurs at too high of a temperature, carbon monoxide (CO) can react with the sulfur to form carbonyl sulfide (COS) via the reaction:

$$CO + \tfrac{1}{n}S_n \rightarrow COS \tag{2}$$

and the product sulfur formed may react further to form SO_2 via the reaction:

$$\tfrac{1}{n}S_n + O_2 \rightarrow SO_2 \tag{3}$$

The objective of this task is to perform a systems analysis of the production of clean syngas, comparing the relative performance and economics of conceptual plant concepts at low and medium temperatures for sulfur recovery, with the medium temperature being the SCOHS process. Parsons has completed a reference Texaco coal gasification plant design case, based on the Tampa Electric Company (TECO) IGCC Demonstration Plant,[3] which will be utilized as the

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[3] "Clean Coal Reference Plants: Integrated Gasification Combined Cycle, Texaco," prepared for the United States Department of Energy, National Energy Technology Laboratory, Contract No. DE-AM26-99T40465, Draft Report, January 2001.

starting point for the analysis. Table 1 indicates the salient features of the baseline plant and the proposed SCOHS process.

Table 1
Plant Features

	Reference	SCOHS Process
Plant Configuration	Texaco IGCC	Texaco IGCC
Acid Gas Removal Temperature	104ºF Amine	275ºF SCOHS
Sulfur Recovery Product	Elemental Sulfur	Elemental Sulfur
Reference for Baseline Cases	CCT Program	T. H. Gardner, NETL
End Product	Power	Power

The SCOHS systems analysis was approached by working with NETL to gain insight into the theory and potential of the process. Todd H. Gardner of the U.S. Department of Energy, National Energy Technology Laboratory, gave the literature review and preliminary recommendation for vessel configuration. This information was used as a basis for developing a preliminary SCOHS acid gas cleanup process design.

1.1 APPROACH

Plant Design Basis – The Texaco reference plant concept was selected as the baseline plant. End product purity requirements will be considered which will determine the level of H_2S to be removed from the syngas. The amine and Claus process for sulfur recovery will be removed from the Texaco plant. Inlet and outlet stream properties will be retained to ensure a consistent basis for process analysis.

SCOHS Design Basis – Parsons will work with NETL to prepare a design basis for the plant that utilizes the SCOHS process. Utilizing input from the NETL review, a preliminary process description and design will be prepared which includes syngas composition and conditions, oxygen air requirements, product yield, and major equipment such as catalytic reactor, condenser, and heat exchangers. Parsons will prepare two variations of the selected baseline SCOHS design, i.e., fixed-bed catalyst and monolithic catalyst, to compare performance and economics with the baseline Texaco plant.

Conceptual Plant Designs – Each plant will produce elemental sulfur as a byproduct. Size and other design features will be retained for each conceptual plant to the extent possible to achieve a fair comparison from plant to plant. Vessels appropriate to the catalysts and reactions will be used in the analysis. For each plant design, the following will be prepared:

- Overall plant description.
- Description of sulfur recovery process.
- Heat and material balance.
- Process flow diagram.
- Capital and operating cost estimate.
- Preliminary economic analysis plant design basis.

1.2 TEXACO REFERENCE PLANT

1.2.1 Overall Plant Description

This IGCC reference plant design is based on the TECO IGCC Demonstration Project, which utilizes an entrained flow oxygen-blown Texaco gasification process. The plant configuration is based on the radiant cooler gasifier mode.

The power generation technology is based on selection of a gas turbine derived from the General Electric 7FA machine. The plant is configured with one gasifier including processes to progressively cool and clean the gas, making it suitable for combustion in the gas turbines (see Figure 1). The resulting plant produces a net output of 287 MWe at a net efficiency of 37.6 percent on an HHV basis. Performance is based on the properties of Pittsburgh No. 8 coal. Overall performance for the entire plant is summarized in Table 2, which includes auxiliary power requirements.

The operation of the combined cycle unit in conjunction with oxygen-blown IGCC technology is projected to result in very low levels of emissions of NOx, SO_2, and particulate (fly ash). A salable byproduct is produced in the form of elemental sulfur. The low level of SO_2 in the plant emissions is achieved by capture of the sulfur in the gas by the amine-based acid gas removal (AGR) process. The AGR process removes approximately 99.5 percent of the sulfur compounds in the fuel gas, resulting in fuel gas with an H_2S concentration of 52 ppmv. The H_2S-rich regeneration gas from the AGR system is fed to a Claus unit with tail gas cleanup.

NOx emissions are limited to approximately 30 ppmv by the use of steam injection. The ammonia is removed with process condensate prior to the low-temperature AGR process. This helps lower NOx levels as well. Selective catalytic reduction (SCR) or selective non-catalytic reduction (SNCR) can reduce emissions further, but is not necessary. Particulate discharge to the atmosphere is limited to low values by the gas washing effect of the syngas scrubber and the AGR absorber.

Figure 1
Block Flow Diagram – Texaco Radiant Cooler IGCC Plant

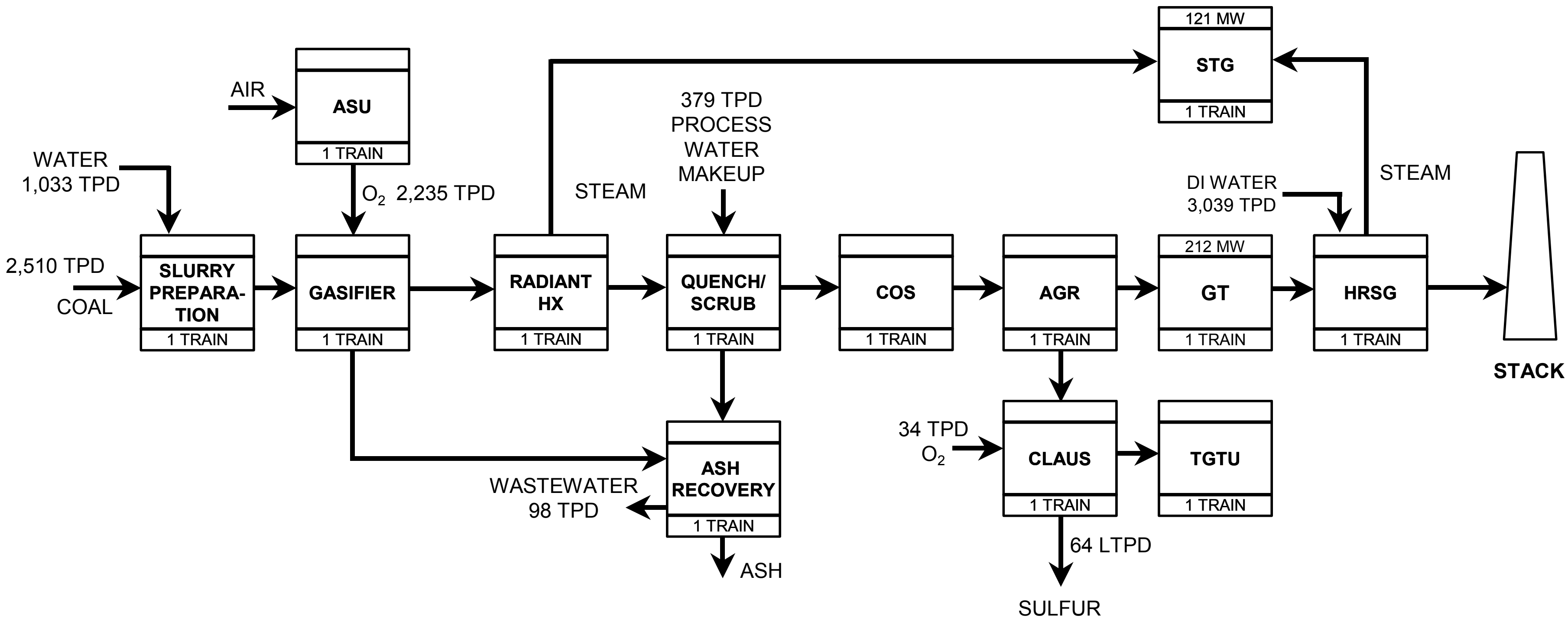

Table 2
TECO IGCC Reference Plant
Plant Performance Summary – 100 Percent Load

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas Turbine	211,570
Steam Turbine	120,260
Generator Losses	(6,210)
Total	**325,620**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	800
Coal Slurry Pumps	290
Condensate Pumps	130
LP/IP BFW Pumps	20
HP BFW Pumps	2,250
Air Separation Plant	19,400
Oxygen Compressor	9,530
Amine Plant	1,170
Claus Plant	200
LP Oxygen Blower	30
Gas Turbine Auxiliaries	600
Steam Turbine Auxiliaries	350
Circulating Water Pumps	1,370
Cooling Tower Fans	800
Slag Handling	130
Transformer Loss	760
Wastewater Treatment	20
Scrubber Pumps	50
Incinerator Blower	50
Miscellaneous Balance of Plant	1,000
TOTAL AUXILIARIES, kWe	**38,950**
Net Power, kWe	286,670
Net Plant Efficiency, % HHV	37.60
Net Heat Rate, Btu/kWh (HHV)	9,085
CONDENSER COOLING DUTY, 10^6 Btu/h	518.0
CONSUMABLES	
As-Received Coal Feed, lb/h	209,196
Thermal Input, kWt	763,301
Total Oxygen (95% pure), lb/h	186,236
Water (for slurry), lb/h	86,066

1.2.2 Reference Plant Acid Gas Removal Processes

1.2.2.1 COS Hydrolysis

Following the syngas scrubber, the gas is reheated to 410°F (210°C) and fed to the COS hydrolysis reactor. The COS is hydrolyzed with steam in the gas, over a catalyst bed to H_2S, which is more easily removed by the AGR solvent. Before the raw synthesis gas can be treated in the sulfur removal process, it must be cooled to 104°F (40°C). During this cooling, part of the water vapor condenses. This water, which contains some NH_3, is sent to the wastewater treatment section. No separate hydrogen cyanide (HCN) removal unit is needed due to the very low HCN concentration in the fuel gas.

1.2.2.2 Acid Gas Removal (AGR)

The promoted monodiethanolamine (MDEA) process was chosen because of its high selectivity toward H_2S and because of the low partial pressure of H_2S in the fuel gas resulting from low gas pressure, necessitating a chemical absorption process rather than a physical absorption process such as the Selexol. The AGR process utilizes an MDEA sorbent and several design features to effectively remove and recover H_2S from the fuel gas stream. The MDEA solution is relatively expensive, and measures are taken to conserve the solution during operations. As the presence of CO causes amine degradation in the form of heat stable salts, an amine reclaimer is included in the process. Also, additional water wash trays are included in the absorber tower to prevent excessive solvent loss due to vaporization.

Fuel gas enters the absorber tower at 104°F (40°C) and 378 psia. Approximately 99.0 percent of the H_2S is removed from the fuel gas stream. The resulting clean fuel gas stream exits the absorber and is heated in a series of regenerative heaters to 310°F (154°C).

The rich MDEA solution is pumped to a regeneration-stripping tower in which the H_2S and CO_2 are stripped from the MDEA by countercurrent contact with CO_2 vapors generated in a steam-heated reboiler. The regenerated H_2S stream contains 79.0 percent CO_2, which can affect the size and efficiency of the Claus reactor. The H_2S stream flows to an H_2S concentration absorber that separates the H_2S from the CO_2. The remaining CO_2-rich stream is incinerated with the vent gas from the tail gas treatment unit. Although not considered in this design, these concentrated streams offer an excellent opportunity for CO_2 capture and sequestration. H_2S is regenerated and sent in a concentrated stream to the Claus plant.

1.2.2.3 Sulfur Recovery System

The sulfur recovery unit is a Claus bypass type sulfur recovery unit utilizing oxygen instead of air and with a Beavon Sulfur Removal (BSR)/Flexsorb tail gas unit. The Claus plant produces molten sulfur by reacting approximately a third of the H_2S in the feed to SO_2, then reacting the H_2S and SO_2 to sulfur and water. The combination of Claus technology and BSR/Flexsorb tail gas technology will result in an overall sulfur recovery exceeding 99.0 percent and a vent gas of less than 50 ppmv of SO_2. Utilizing oxygen instead of air in the Claus plant reduces the overall cost of the sulfur recovery plant. The sulfur plant will produce approximately 65 tons per day. Feed for this case consists of acid gas from both acid gas cleanup units and a vent stream from the sour water stream in the gasifier section. Vent gas from the tail gas unit will be vented to the incinerator, and the resulting vent will meet the air quality standards of 50 ppmv of SO_2.

1.2.2.4 Sour Gas Stripper

The sour gas stripper removes ammonia (NH_3), SO_2, and other impurities from the waste stream of the scrubber. The sour gas stripper consists of a sour drum that accumulates sour water from the gas scrubber and condensate from syngas coolers. Sour water from the drum flows to the sour stripper, which consists of a packed column with a steam-heated reboiler. Sour gas is stripped from the liquid and sent to the sulfur recovery unit. Remaining water is sent to wastewater treatment.

1.3 RETROFITTING THE REFERENCE PLANT WITH SCOHS

Raw syngas exiting the gasifier goes through the first syngas cooler to raise HP saturated steam, followed by a series of coolers and a syngas scrubber. The COS process will be retained, but will be modified to operate with exit gas at 402 psia and 275°F. At this point, the SCOHS process will be installed. Gas exiting the SCOHS will actually be cleaner than the gas leaving the MDEA unit, and will be maintained at 362.5 psia and 310°F going into the gas turbine. The fuel gas will contain essentially zero H_2S and about 20 ppmv COS. The gas temperature leaving the SCOHS may be slightly higher than 310°F to retain sensible heat going to the gas turbine.

The following equipment will be removed from the TECO plant:

- Sour gas coolers and knockout drum.
- MDEA unit and clean gas reheat heat exchanger.
- Claus plant.
- Tail gas treatment unit and tail gas incinerator.

2. EXPERIENCE WITH A PROCESS SIMILAR TO SCOHS

The activated-carbon process[4], which was developed by I. G. Farbenindustrie during the 1920s, takes advantage of the catalytic action of activated carbon in promoting the oxidation of H_2S to elemental sulfur at ambient temperatures. The sulfur deposited on the activated carbon was recovered by extraction with an appropriate solvent, ammonium sulfide, and the carbon was reused until attrition of the carbon particles became excessive. The activated-carbon process has the distinct advantage that very pure sulfur was obtained with a relatively simple operation.

While similar to the SCOHS process, this process differs in two fundamental ways:

- The activated-carbon process operated at low temperature and pressure, while the SCOHS process will operate at high pressure and temperatures above the dewpoint of the coal-derived synthesis gas.
- The activated-carbon process was regenerated by the use of solvent extraction. Since activated carbon contains significant microporosity, complete regeneration with a liquid solvent would take several steps. Additionally, the solvent extraction step must be followed by an energy-intensive evaporation step to recover the solvent. In contrast, the SCOHS process will regenerate with heated nitrogen that is readily available from the oxygen separation plant. Here, full regeneration is possible since nitrogen can easily penetrate into the pores of the catalyst, and utilization of an energy-intensive solvent extraction step is mitigated since the solvent is a gas rather than a liquid.
- The SCOHS process may employ catalysts other than activated carbon. Many metal oxides, binary metal oxide, zeolites, and metal carbides have demonstrated some level of activity for reaction (1).

2.1 PROCESS DESCRIPTION[5,6]

The sour gas is passed to a carbon bed, after addition of air and a small amount of ammonia. In order to ensure complete reaction, it is customary to add approximately 50 percent more air than is stoichiometrically required. The ammonia, which increases the rate of oxidation quite appreciably, is added in the proportion of 5 volumes of ammonia to 100 volumes of H_2S[7]. When the bed is saturated, as evidenced by the appearance of small amounts of H_2S in the treated gas, the gas flow is switched to a second bed, and the first bed is regenerated.

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[4] Kohl, A. L. and F. C. Riesenfeld, "Gas Purification, Third Edition," Gulf Publishing Company, Houston, Texas, 1979, 402-406.

[5] Engelhardt, A. 1928. "Gas-u. Wasserfach" 71(13):290.

[6] Kronacher, H. K. 1931. "Gas Age-Record" 68(2):37.

[7] Francis, W. 1951. "Engineering" 172(Aug. 10): 180.

Regeneration is carried out by extraction of the sulfur in several successive stages with a 15 percent aqueous ammonium sulfide solution, followed by steaming of the bed for the removal of residual ammonium sulfide. Solution is first pumped into the saturated bed until the carbon layer is completely covered with liquid. A few minutes are allowed for dissolving the sulfur, and the solution is then drained back. This treatment is repeated with the solutions from different tanks so that the last solution contacts essentially sulfur-free carbon. The carbon, which contains practically sulfur-free ammonium sulfide solution, is now treated with saturated steam at 212°F and is then ready for further service. The vapors from the steam treatment, which contain ammonia, H_2S, and water, are condensed in a spray condenser, and the condensate is accumulated in another tank, whence it is used for regeneration.

After the extraction process has been repeated several times, the solution becomes saturated; this is indicated by a sulfur content of about 1.7 to 2.5 lb/gal of liquid. The saturated solution is pumped to another tank, from which it flows by gravity to the evaporator. Here the solution is heated by addition of steam and polysulfides are decomposed. The overhead vapors, containing H_2S, ammonia, and water, are condensed in the condenser and solid sulfur with some water removed from the bottom of the evaporator. The water, which is separated from the sulfur in a centrifuge, is sprayed into the condenser. The sulfur obtained is granular and contains 1 to 2 percent moisture. By operating at pressures of 1.5 to 2 atmospheres and correspondingly higher temperatures, liquid sulfur can be withdrawn from the bottom of the evaporator.

2.2 DESIGN AND OPERATION

Space velocities of 350 to 400 volumes of gas per hour per volume of carbon are customary. A typical activated-carbon purifier, as used in Germany, consists of a cylindrical carbon steel vessel, 13 feet in diameter, in which the carbon is placed on a horizontal grid to a depth of approximately 4 feet. Such a unit is capable of processing approximately 200,000 cubic feet of gas/hour with a pressure drop of about 25 inches of water column[8]. The activated carbon normally used in Germany retains approximately 25 to 35 pounds of sulfur per cubic foot at saturation.

In order to accomplish complete H_2S removal without COS formation, the temperature of the bed had to be maintained below 140°F. Because of the high heat of reaction, gases containing more than 400 grains H_2S per 100 cubic feet cannot be treated satisfactorily by cooling the bed. The rate of oxidation is influenced favorably by the presence of water vapor in the gas, and it is therefore advantageous to saturate the gas with water before it contacts the activated carbon. As already mentioned, small amounts of ammonia increase the catalytic activity of the carbon considerably.

The activated carbon used in the I.G. Farbenindustrie process for H_2S removal is prepared from low-temperature, brown-coal coke, which is ground to a particle size of 1 to 4 mm in diameter.

[8] Spichal, W. 1953. "Gas-u. Wasserfach" 94(23):674-684.

The carbon is activated by heating with combustion gases and steam at approximately 1500°F for several hours. The resulting product has a bulk density of about 25 pounds per cubic foot and is capable of absorbing sulfur to the extent of 100 to 150 percent of its weight. A good grade of activated carbon will withstand 20 to 30 cycles of saturation and regeneration.

Table 3 summarizes the design basis for the activated-carbon process as described above.

Table 3
Activated-Carbon Process
Design Basis

	Activated-Carbon Process
Syngas (lb mol/h)	539
Syngas (lb/h)	11,162
Molecular weight	20.7
Temperature (°F)	125
Pressure (psia)	17
Sulfur concentration (ppm)	3,200
Air stoichiometry	2
Air (scf/h)	929
Catalyst field packed density (lb/cf)	25
Catalyst loading (lb sulfur/lb catalyst)	150%
Space velocity (h^{-1})	400
Syngas and air (scf/h)	194,523
Syngas and air (acf/h)	200,000
Absorbent catalyst (scf)	486
Cycle time absorption and regeneration (h)	660.5
Superficial velocity (ft/sec)	0.46
Number of vessel pairs or vessels for monolith	1
Vessel ID (ft)	12.4
Bed height (ft)	4.0
Catalyst life (cycles)	30
Catalyst replacement time (month)	27.1
Catalyst replacement rate (ton/y)	4
Initial catalyst charge (ton)	12

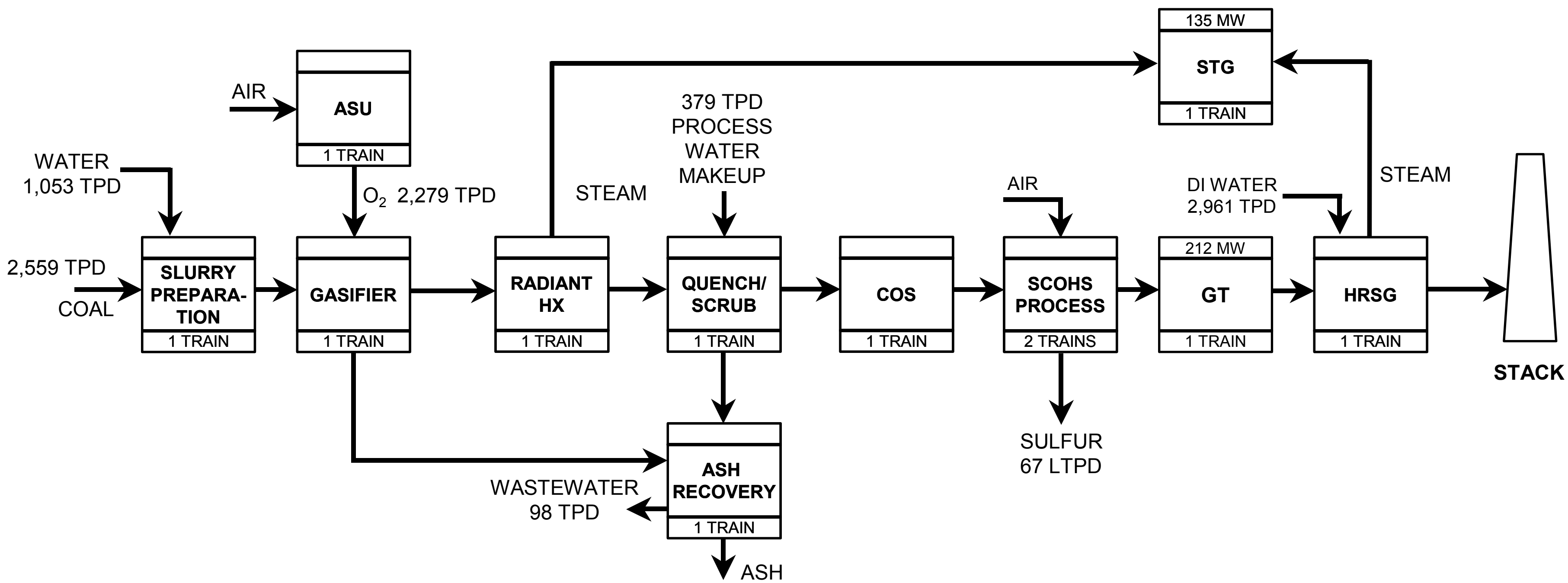
3. SCOHS CASE 1 – FIXED-BED CATALYST – DESIGN

Table 4 summarizes the major design parameters for the SCOHS process resulting from the experimental data generated to date. Because of the relatively high space velocity and low catalyst loading, the cycle time is short and the catalyst replacement rate, based on 100 cycles, is high. Table 4 also presents the recommended design parameters for this SCOHS study. Figure 2 is the block flow diagram of the IGCC plant with the SCOHS process replacing the conventional sulfur removal processes.

Table 4
SCOHS Process
Design Basis

	SCOHS Experimental Data	SCOHS Design Basis Fixed Bed
Syngas (lb mol/h)	21,933	21,933
Syngas (lb/h)	452,117	452,117
Molecular weight	20.6	20.6
Temperature (°F)	275	275
Pressure (psia)	402	402
Sulfur concentration (ppm)	8,441	8,441
Air stoichiometry	4	4
Air (scf/h)	265,856	265,856
Catalyst field packed density (ln/cf)	36	36.0
Catalyst loading (lb sulfur/lb catalyst)	10%	50%
Space velocity (h^{-1})	2,500	1,000
Syngas and air (scf/h)	8,139,803	8,139,803
Syngas and air (acf/h)	444,659	444,659
Absorbent catalyst (scf)	3,256	8,140
Cycle time absorption and regeneration (h)	4.0	49.5
Superficial velocity (ft/sec)	1.00	0.50
Number of vessel pairs or vessels for monolith	1	2
Vessel ID (ft)	12.5	12.5
Bed height (ft)	26.4	33.0
Catalyst life (cycles)	100	100
Catalyst replacement time (month)	0.5	6.8
Catalyst replacement rate (ton/y)	2,076	415
Initial catalyst charge (ton)	117	293

Figure 2
Block Flow Diagram – Texaco Radiant Cooler
IGCC Plant with Fixed-Bed SCOHS Process



Reducing the space velocity increases the amount of catalyst, but not the catalyst makeup rate, and increases the catalyst loading, thus resulting in longer cycle times. By reducing the space velocity in the SCOHS process to 1,000, it was assumed that the catalyst loading would increase to 50 percent. This is not inconsistent with the activated-carbon process, where a space velocity of 400 resulted in a catalyst loading of 150 percent. This would result in a cycle time of 50 hours, a catalyst replacement time of 6.8 months based on 100 cycles, and a sorbent makeup rate of 415 tons per year. Four vessels, each with a 12.5-foot ID and a sorbent bed height of 33 feet, would be required.

3.1 PLANT DESCRIPTION

The process and equipment bases for the absorbent/regenerator design were established for the SCOHS process using the existing database, where possible, and reasonable engineering judgments. Since the SCOHS process is still in the development stage, design assumptions were made that have not been verified by large-scale, long-term testing. Sensitivity and trade-off studies evaluated the impact of these assumptions on costs.

3.1.1 Absorption

For the fixed-bed cases, desulfurization of the fuel gas is accomplished in a fixed bed of granular activated-carbon catalyst. The activated-carbon catalyst is in the form of 1/8- to 3/16-inch-diameter granules. The properties of the granules assumed for this study are given in Table 5. Based on experimental data from Todd H. Gardner at the U. S. Department of Energy, National Energy Technology Laboratory, it was assumed that all SCOHS reactors have only 20 ppmv COS in exit, independent of inlet H_2S. Therefore, with 8,441 ppmv H_2S inlet, 20 ppmv COS in outlet results in 99.76 percent recovery.

Space velocity, defined as the ratio of gas flow rate at standard conditions to sorbent bulk volume, is proportional to reciprocal gas residence time in the bed. A higher space velocity decreases the amount of catalyst necessary and thus capital costs. However, the cycle time would decrease, resulting in more cycles per year of plant operation. We have assumed a space velocity of 1,000 hr^{-1}, which falls in the range of the NETL tests and the operating experience of the activated-carbon process.

The catalyst loading and gas space velocity determine the initial time to breakthrough for a fuel gas with a specified sulfur concentration. For this study, it was assumed that the catalyst loading was 50 percent at a space velocity of 1,000 hr^{-1} (STP). This results in a 24-hour absorption cycle and, assuming a catalyst life of 100 cycles, a catalyst replacement time of 6.6 months.

Since the sulfur reactions, which occur during absorption, are exothermic, bed temperature is controlled to a maximum of 310°F (the turbine inlet temperature) by maintaining the syngas temperature at 275°F.

Table 5
Granular Activated-Carbon Specifications

Specifications	
Peroxide No.*	14 max
Iodine No., mg/g	800 min
Butane activity, wt%	15.6 min
Ash, by weight%	7 max
Moisture, by weight%, as packed	2 max
Hardness No.	97 min
Apparent density, g/cc	0.56 min
Mean particle diameter, mm	3.7 min
U.S. sieve series:	
Percent on 4 mesh	15% max
Percent through 7 mesh	8.0% max

* Peroxide number utilizes the rate of decomposition of hydrogen peroxide by the carbon and is an indicator of the amount of catalytic activity.

3.1.2 Bed Regeneration and Sulfur Recovery

For the SCOHS baseline design, regeneration of the catalyst bed is required when the catalyst reaches 50 percent of its weight with sulfur. The method selected for catalyst regeneration includes heating the bed and extracting sulfur vapors with hot circulating nitrogen. In order to regenerate the catalyst, it must be heated to a temperature at which sulfur vapor pressure increases and the vapors can be swept away. Figure 3 shows the schematic block flow diagram for the regeneration step. By appropriate valving, a continuous nitrogen loop sends nitrogen through the bed, through a regenerative heat exchanger, and through a sulfur condenser. Following the sulfur condenser, liquid sulfur can be removed from the gas/liquid separator. The nitrogen is then recirculated with a boost compressor.

The closed-loop nitrogen system is designed to achieve this. After heating the catalyst bed, sulfur vapor is swept up by the circulating nitrogen and condensed as a liquid in the sulfur condenser. The nitrogen is then reheated and recirculated until all the sulfur is recovered. The nitrogen is then cooled and used to cool the bed to the initial absorber operating temperature. Regeneration occurs in a series of stages that occur in a 24-hour window, followed by absorbing for the next 24 hours. By using the regenerative heat exchanger and a fired heater, nitrogen temperature can be controlled through the various stages of the regeneration cycle.

The vapor pressure of elemental sulfur as a function of temperature is shown on Figure 4. It was determined that the best temperature for regeneration is 650°F, at which point the vapor pressure of sulfur is 2.8 psia. This is a reasonable temperature for control and material selection, and will permit full regeneration in the available time frame.

Figure 3
Flow Diagram – Fixed-Bed SCOHS Retrofit

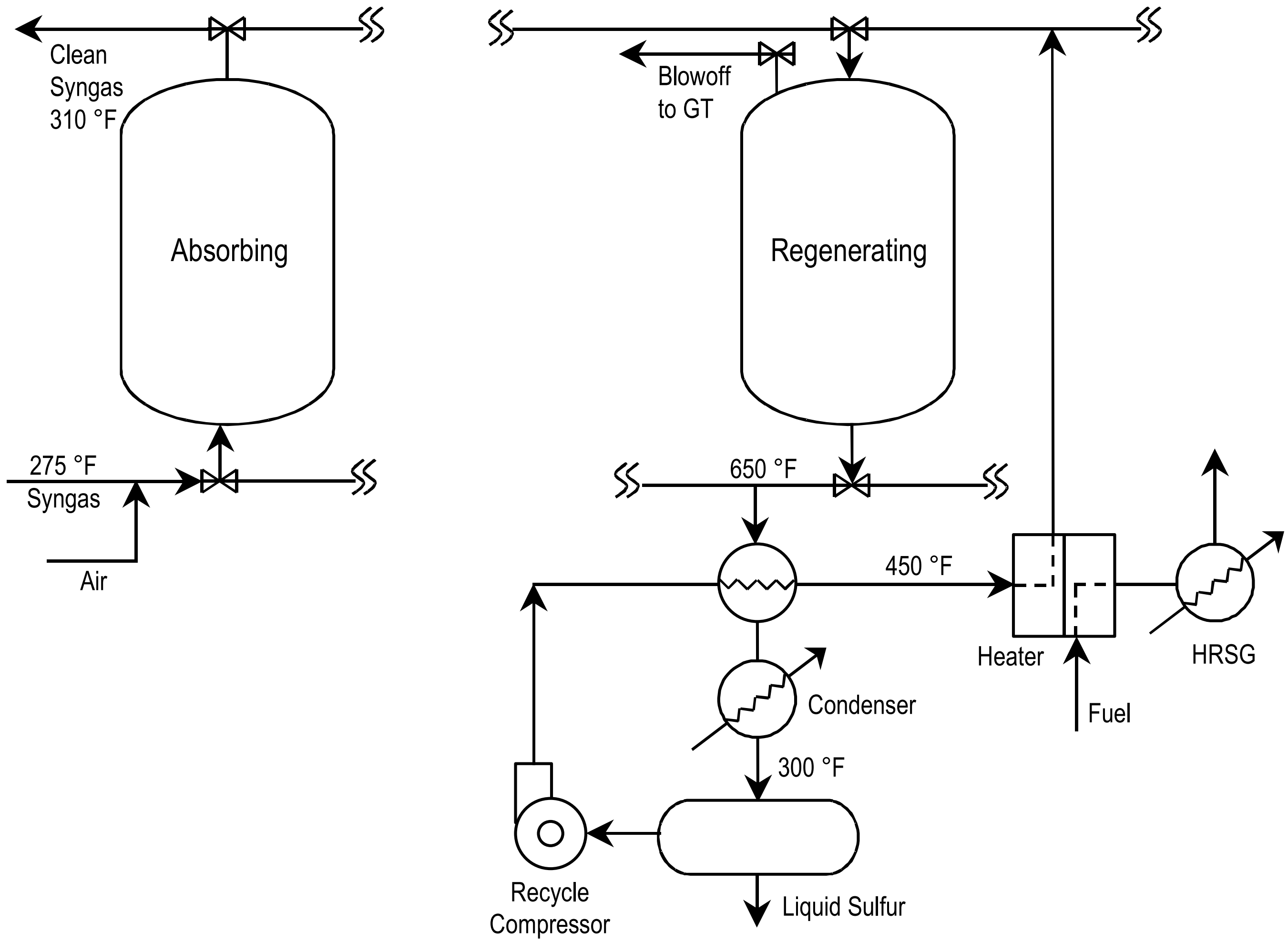


Figure 4
Sulfur Vapor Pressure

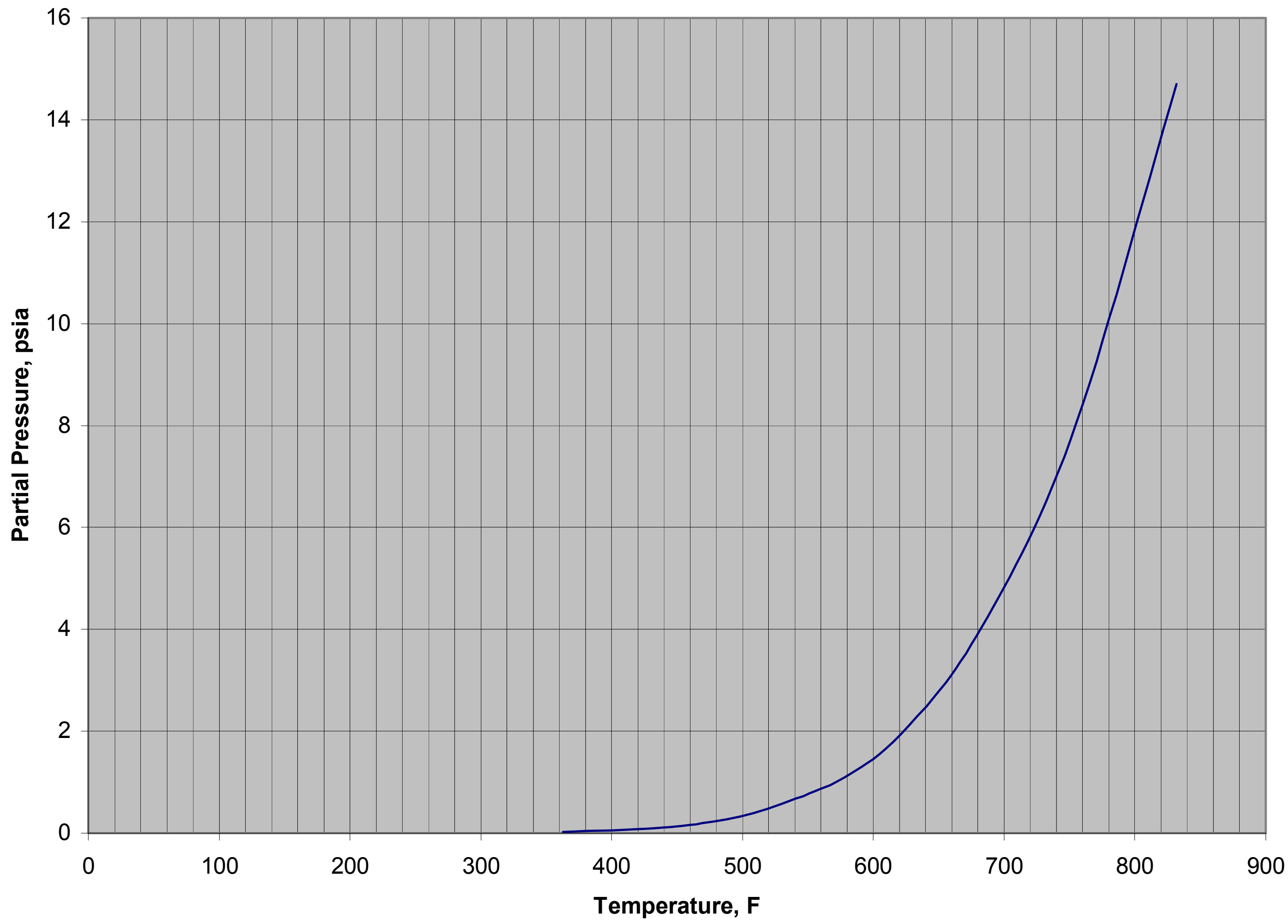
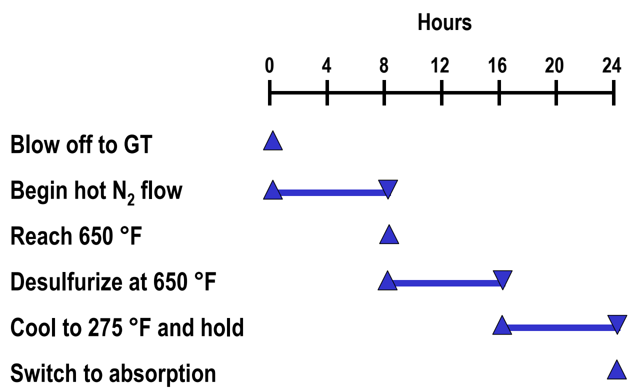


Figure 5 is a time sequence, which indicates the cycles required for bed regeneration. Initially, the bed is isolated and depressurized. The syngas contained in the vessel is released to the gas turbine or a convenient burner elsewhere in the plant. Each regeneration gives up 34,000 scf of syngas or 8.5 MMBtu of fuel value.

Figure 5
Regeneration Sequence



The depressurized vessel is then filled with 60 psia nitrogen and integrated with the nitrogen loop. It is heated by circulating nitrogen, which is heated from a fired heater. Some cooling is needed toward the end of the cycle to keep the blower temperature below 400°F. The regeneration parameters are shown in Table 6.

Table 6
Regeneration Parameters

Initial bed temperature	315°F
Final bed temperature	650°F
Catalyst bed mass	141,500 lb
Heat flux into bed	1.546 MMBtu/h
Time to heat bed to final temperature	8.55 hours
Nitrogen temperature	750°F in, 650°F out
Nitrogen flow rate through bed	60,000 lb/h, (12,700 scfm)
Vessel space velocity	180 hours^{-1}
Superficial velocity at 650°F and 60 psia	0.95 ft/sec

When the bed reaches 650°F, the nitrogen circulation continues. The sulfur condenser is utilized to cool the gas to 300°F and condense sulfur vapors from the nitrogen. Liquid sulfur is separated from the gas and the gas is recycled. With regenerative heat exchange, the additional heat to maintain the bed at 650°F is significantly reduced. The sulfur recovery parameters are shown in Table 7.

Table 7
Sulfur Recovery Parameters

Regeneration temperature	650°F
Sulfur vapor pressure	2.8 psia
Maximum sulfur recovery rate	12,224 lb/h
Minimum time to fully regenerate	5.8 hours
Recommended regeneration time	8 hours

Following catalyst regeneration, the bed is allowed to cool back to 275°F by continuing to circulate nitrogen and extract heat using the sulfur condenser. The sulfur condenser has a capacity of 3.12 MMBtu/hour, which results in bed cooling to 275°F in 5 hours.

The cool catalyst bed, free of sulfur, is maintained at 275°F for approximately 1 hour. It is then switched to the active syngas stream for the next absorbing cycle, and the parallel catalyst vessel is taken off stream to be regenerated.

3.1.3 Heat and Material Balance

The fixed-bed SCOHS retrofit of the Tampa Electric IGCC Demonstration Project results in a plant producing a net output of 301 MWe at a net efficiency of 38.7 percent on an HHV basis. Performance is based on the properties of Pittsburgh No. 8 coal. Overall performance for the entire plant is summarized in Table 8, which includes auxiliary power requirements.

Table 8
Tampa Electric IGCC Reference Plant with Fixed-Bed SCOHS Retrofit
Plant Performance Summary – 100 Percent Load

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas turbine	211,650
Steam turbine	135,150
Generator losses	(6,460)
Total	**340,340**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	820
Coal Slurry Pumps	300
Condensate Pumps	140
LP/IP BFW Pumps	30
HP BFW Pumps	2,240
Air Separation Plant	19,280
Oxygen Compressor	9,720
SCOHS Plant Auxiliaries	1,290
Gas Turbine Auxiliaries	600
Steam Turbine Auxiliaries	350
Circulating Water Pumps	1,690
Cooling Tower Fans	990
Slag Handling	130
Transformer Loss	800
Wastewater Treatment	20
Scrubber Pumps	70
Miscellaneous Balance of Plant	1,000
TOTAL AUXILIARIES, kWe	**39,460**
Net Power, kWe	300,880
Net Plant Efficiency, % HHV	38.7
Net Heat Rate, Btu/kWh (HHV)	8,828
CONDENSER COOLING DUTY, 10^6 Btu/h	640.1
CONSUMABLES	
As-Received Coal Feed, lb/h	213,282
Thermal Input, kWt	778,210
Total Oxygen (95% pure), lb/h	189,884
Water (for slurry), lb/h	87,747

3.1.4 Process Flow Diagram

Figure 6 is the process flow diagram resulting from the fixed-bed SCOHS heat and material balance.

Figure 6 (3 pages) follows.

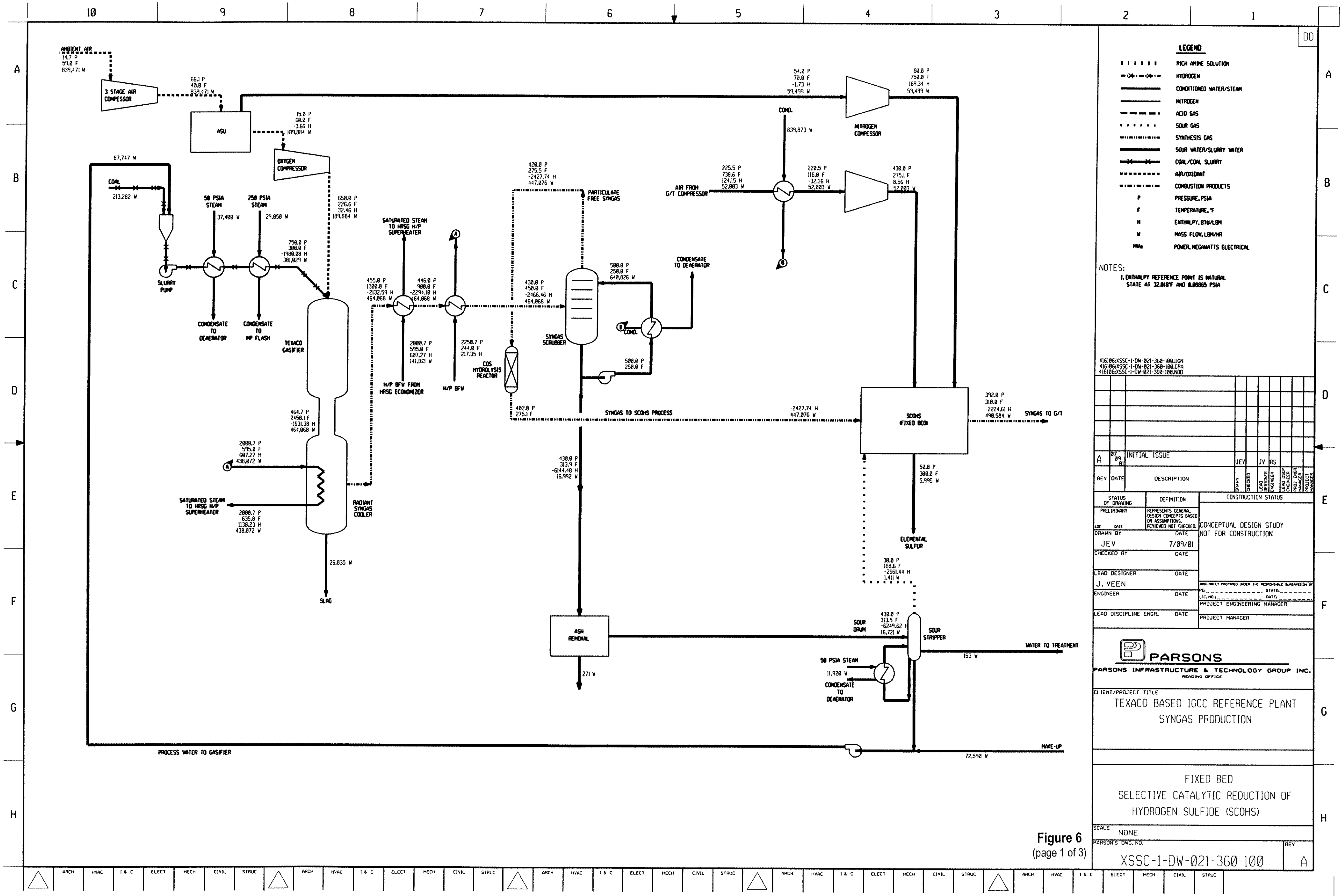
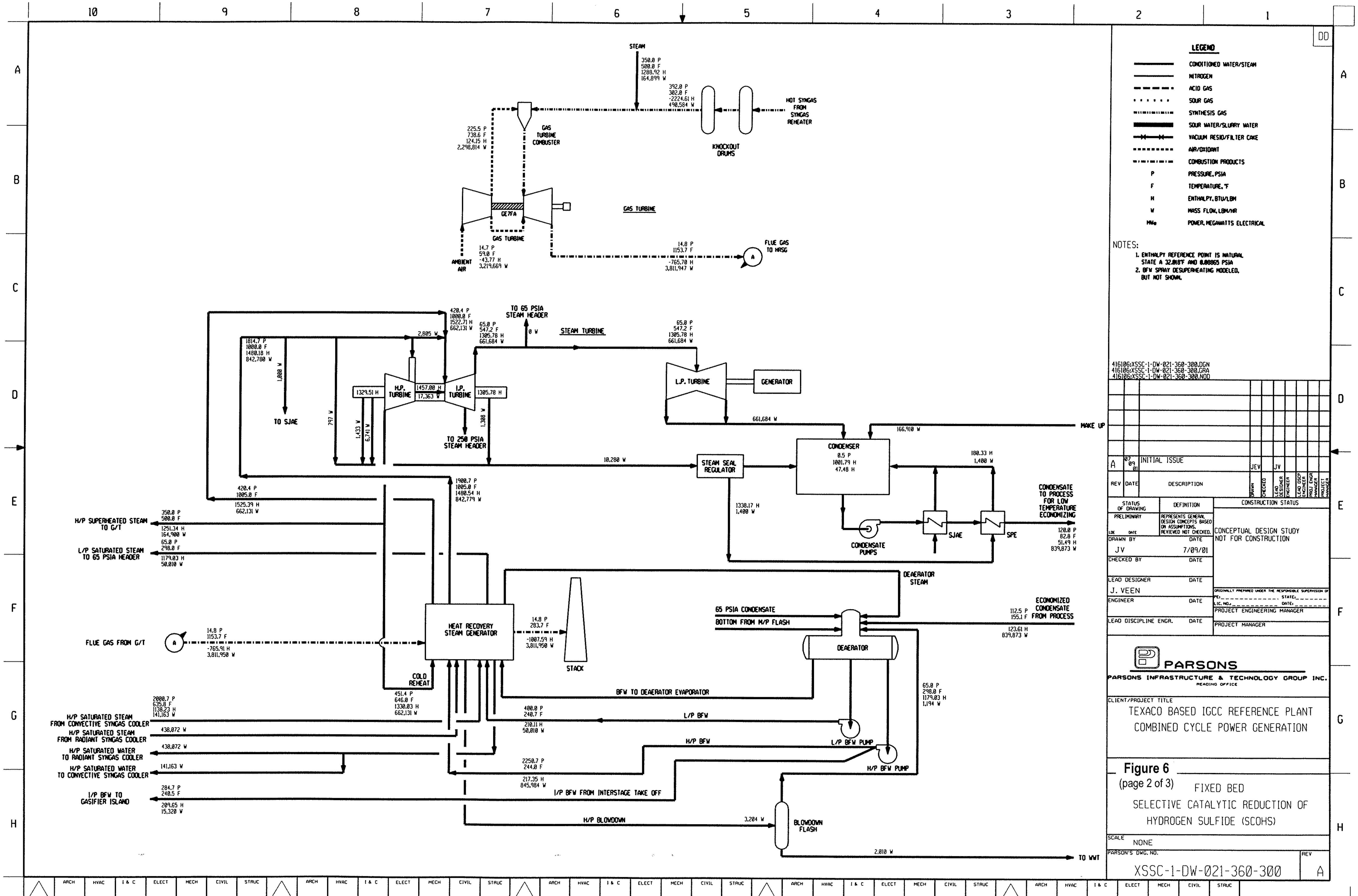


Figure 6
(page 1 of 3)


LEGEND
CONDITIONED WATER/STEAM
NITROGEN
ACID GAS
SOUR GAS
SYNTHESIS GAS
SOUR WATER/SLURRY WATER
VACUUM RESID/FILTER CAKE
AIR/OXIDANT
COMBUSTION PRODUCTS
P PRESSURE, PSIA
F TEMPERATURE, °F
H ENTHALPY, BTU/LBM
W MASS FLOW, LBM/HR
MWe POWER, MEGAWATTS ELECTRICAL
NOTES:
1. ENTHALPY REFERENCE POINT IS NATURAL STATE A 32.018°F AND 0.08865 PSIA
2. BFW SPRAY DESUPERHEATING MODELED, BUT NOT SHOWN.
STEAM
GAS TURBINE COMBUSTER
KNOCKOUT DRUMS
HOT SYNGAS FROM SYNGAS REHEATER
GE7FA
GAS TURBINE
AMBIENT AIR
FLUE GAS TO HRSG
TO 65 PSIA STEAM HEADER
STEAM TURBINE
H.P. TURBINE
I.P. TURBINE
L.P. TURBINE
GENERATOR
TO SJAE
TO 250 PSIA STEAM HEADER
MAKE UP
CONDENSER
STEAM SEAL REGULATOR
CONDENSATE PUMPS
SJAE
SPE
CONDENSATE TO PROCESS FOR LOW TEMPERATURE ECONOMIZING
H/P SUPERHEATED STEAM TO G/T
L/P SATURATED STEAM TO 65 PSIA HEADER
DEAERATOR STEAM
65 PSIA CONDENSATE
BOTTOM FROM H/P FLASH
ECONOMIZED CONDENSATE FROM PROCESS
DEAERATOR
FLUE GAS FROM G/T
HEAT RECOVERY STEAM GENERATOR
STACK
COLD REHEAT
BFW TO DEAERATOR EVAPORATOR
H/P SATURATED STEAM FROM CONVECTIVE SYNGAS COOLER
H/P SATURATED STEAM FROM RADIANT SYNGAS COOLER
H/P SATURATED WATER TO RADIANT SYNGAS COOLER
H/P SATURATED WATER TO CONVECTIVE SYNGAS COOLER
I/P BFW TO GASIFIER ISLAND
L/P BFW
H/P BFW
L/P BFW PUMP
H/P BFW PUMP
I/P BFW FROM INTERSTAGE TAKE OFF
H/P BLOWDOWN
BLOWDOWN FLASH
TO WWT
416106:XSSC-1-DW-021-360-300.DGN
A 07 09 01 INITIAL ISSUE JEV JV
REV DATE DESCRIPTION
STATUS OF DRAWING
PRELIMINARY
CONSTRUCTION STATUS
CONCEPTUAL DESIGN STUDY NOT FOR CONSTRUCTION
DRAWN BY JV DATE 7/09/01
CHECKED BY
LEAD DESIGNER J. VEEN
ENGINEER
LEAD DISCIPLINE ENGR.
PROJECT ENGINEERING MANAGER
PROJECT MANAGER
PARSONS
PARSONS INFRASTRUCTURE & TECHNOLOGY GROUP INC.
READING OFFICE
CLIENT/PROJECT TITLE
TEXACO BASED IGCC REFERENCE PLANT
COMBINED CYCLE POWER GENERATION
Figure 6
(page 2 of 3)
FIXED BED
SELECTIVE CATALYTIC REDUCTION OF
HYDROGEN SULFIDE (SCOHS)
SCALE NONE
PARSON'S DWG. NO.
XSSC-1-DW-021-360-300
REV A

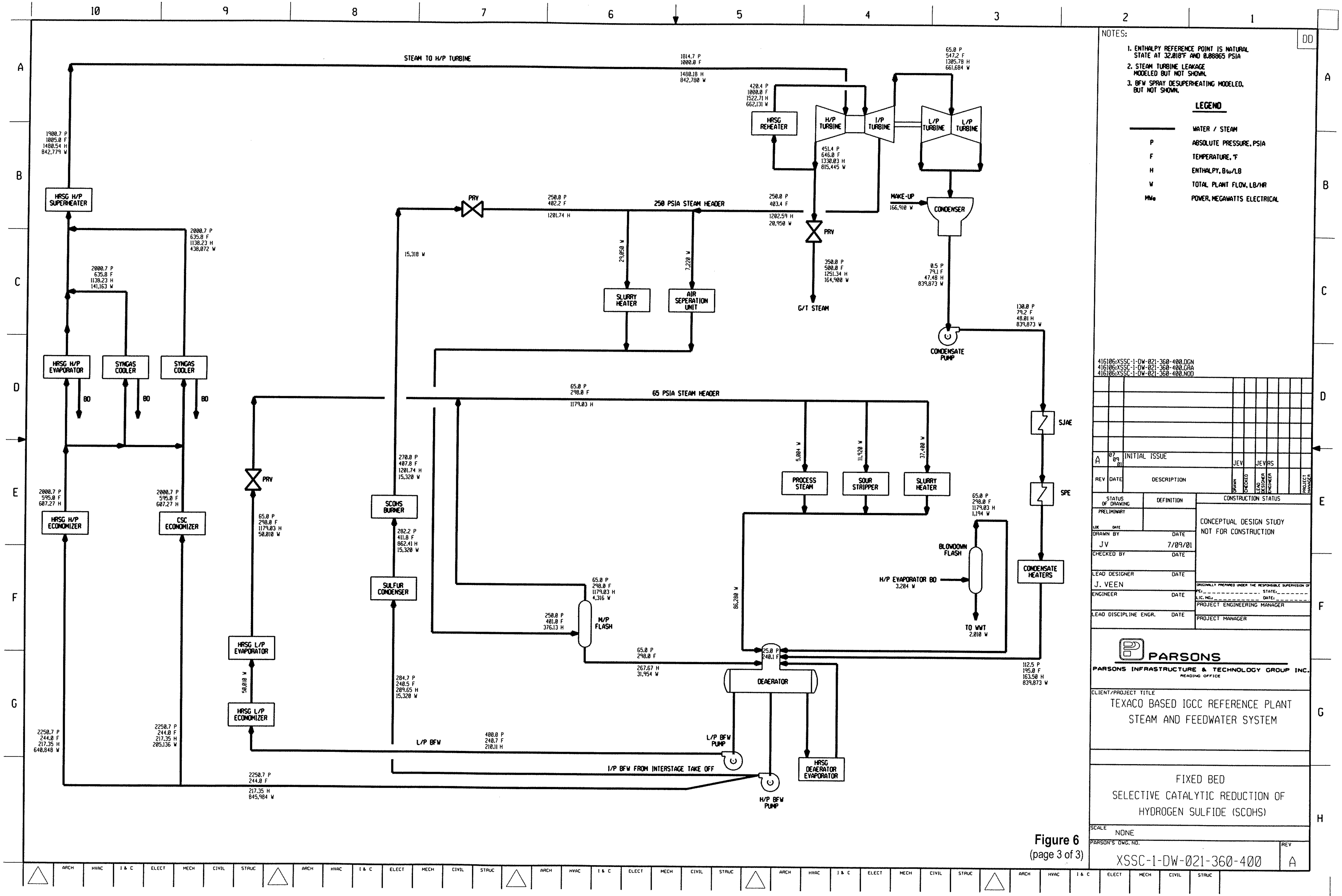


Figure 6
(page 3 of 3)

3.1.5 Equipment Design

The approach taken in this study was to assume multiple, low-profile vessels for absorption and regeneration. This approach has many advantages including:

- Valves and manifolds can be located at grade with a minimum of hot gas piping.
- The low profile provides easy access to the vessels, simplifying sorbent change out and maintenance.

In general, dimensions of the vessels needed to contain the required volume of catalyst were established on the basis of construction costs and operability. For a given gas flow and catalyst bed volume, a deep bed is more effective than a shallow bed, in that it permits the catalyst to attain a higher average loading. This advantage is gained at the expense of pressure drop since the deep bed must be operated at a higher gas velocity. Allowable gas velocities are also limited by considerations of particle entrainment and bed agitation as well as pressure drop. In general these constraints result in a height/diameter ratio of less than 3:1 and velocities of less than 1 ft/sec.

For the fixed bed, the two major equipment components are the absorber vessels, with associated piping, headers, valves, and the regeneration gas compressors and heaters. The four vessels are vertical with a carbon steel shell. Four valves per vessel are necessary to regulate the absorption and regeneration cycles.

3.2 COST ANALYSIS

3.2.1 Capital and Operating Cost Estimate

The fixed-bed SCOHS retrofit of the Texaco-based IGCC Reference Plant resulted in a plant that is close in performance and size to the original plant. The overall capital cost difference between the plants is small, notwithstanding the significant cost changes in the sulfur removal and sulfur recovery processes. Table 9 was prepared to show the factored adjustment of capital from the Texaco-based IGCC Reference Plant to the plant with fixed-bed SCOHS retrofit. The initial estimate for the cost of the SCOHS equipment is nominally $10 million. Although the absolute plant total capital requirement changed by only 2 percent, the SCOHS capital cost per kW is lowered by 6 percent because of higher plant efficiency and higher power production.

Table 9
Capital Cost Changes with Fixed-Bed SCOHS Retrofit

		Texaco IGCC Plant Size 287 MWe				Fixed Bed SCOHS Plant Size 301 MWe			
No.	**Account**	**Basis**	**Unit**	**TPC 1,000$**	**$/kW**	**Basis**	**Unit**	**TPC 1,000$**	**$/kW**
1	Coal Handling	209,196	lb/h	$14,931	$52	213,282	lb/h	$15,223	$51
2	Coal Preparation	209,196	lb/h	$14,427	$50	213,282	lb/h	$14,709	$49
3	Feedwater Pumps	2,250	kW	$14,883	$52	2,240	kW	$14,817	$49
4	Gasifier	1	Train	$50,251	$175	1	Train	$50,251	$167
4.3	ASU	189,106	lb/h	$39,797	$139	189,885	lb/h	$39,961	$133
5	Gas Cleanup	MDEA/Claus/TGTU		$23,751	$83	Fixed Bed SCOHS		$10,000	$33
6	Combustion Turbine	1	7FA	$54,332	$190	1	7FA	$54,332	$181
7	HRSG	1	Each	$21,878	$76	1	Each	$21,878	$73
8	Steam Turbine	120	MW	$27,660	$96	135	MW	$31,085	$105
9	Cooling Water	518	10^6 Btu	$15,090	$53	640	10^6 Btu	$18,647	$62
	BOP[(1)] Subtotal	--	--	$65,572	$229	--	--	$65,572	$218
	Total Plant Costs			**$342,572**	**$1,195**			**$336,474**	**$1,118**
	Others			**$37,353**	**$130**			**$36,353**	**$121**
	Total Capital Requirements			**$379,925**	**$1,325**			**$372,827**	**$1,239**

(1) BOP includes ash, accessory electrical, I&C, site, and building systems.

3.2.2 Preliminary Economic Analysis

The cost of electricity (COE) for the fixed-bed SCOHS retrofit was determined by adjusting the COE from the Texaco-based IGCC Reference Plant. Table 10 shows the changes in the components making up the COE. The COE is based on a fuel cost of $1.25 per MMBtu and an annual plant capacity factor of 80 percent. The change in COE with the fixed-bed SCOHS retrofit amounts to a reduction of 6.4 percent.

Table 10
Change in Cost of Electricity with Fixed-Bed SCOHS Retrofit

	Texaco IGCC Plant Size 287 MWe		Fixed Bed SCOHS Plant Size 301 MWe	
	1,000$/y	**$/kW-y**	**1,000$/y**	**$/kW-y**
Capital charge	$52,432	$182.90	$51,450	$171.00
O&M	$12,126	$42.30	$11,626	$38.64
Consumables	$2,332	$8.13	$1,832	$6.09
Sulfur credit @ $47/LT	($898)	($3.13)	($915)	($3.04)
Fuel @ $1.25/ MMBtu	$22,634	$78.95	$23,081	$76.71
Total	**$88,626**	**$309.16**	**$87,074**	**$287.56**
COE @ 80% CF $/MWh	**44.11**		**41.30**	

4. SCOHS CASE 2 – MONOLITHIC CATALYST BED – DESIGN

4.1 PLANT DESCRIPTION

Figure 7 is the block flow diagram of the IGCC plant with the monolith SCOHS process replacing the conventional sulfur removal processes. Figure 8 is a process flow diagram for the SCOHS monolith reactor.

4.1.1 Process Design

The monolith catalyst version of the SCOHS process utilizes a carbon-fiber-based porous carbon monolithic catalyst (similar to that developed at the Oak Ridge National Laboratory)[9] to continuously catalyze the SCOHS reaction and produce liquid sulfur without a regeneration step.

The material is manufactured from milled carbon fibers and powdered phenolic resin, slurried in water, and vacuum molded. The monolith is carbonized at 650°C. The composite is strong and porous, allowing fluids to flow easily through the material. The unique feature of the composite structure is the surface area decrease as a function of steam activation at 850°C. With limited activation, the mesopore structure surface area is retained at 300 to 500 m^2/gram. With increased burnoff, the surface area drops to 200 m^2/gram, which, coincidentally, is about equal to the surface area of the Calgon carbon used in the fixed-bed concept. The surface area appears to be constant with the amount of burnoff, indicating that the carbon fibers are non-porous. With the open area of the concept and limited porosity, the SCOHS reaction should have adequate surface, but the liquid sulfur product should release from the surface and flow through the monolith bed.

The SCOHS reaction still requires two reactor trains but, since offline regeneration is not necessary, only two vessels instead of four are required. Because of the low bulk density of the monolith composite, only about half as much catalyst by weight is required. Compressed air at 400 percent stoichiometric of the H_2S to sulfur oxidation is mixed with the fuel gas leaving the COS reactor at 402 psia and 275°F.

The gas enters the reactor and the H_2S and oxygen react to form liquid elemental sulfur. The sulfur is carried through the composite monolith by gas flow and gravity to a sump at the bottom of the reactor vessel. The exothermic sulfur reaction causes the gas temperature to increase to 315°F. A high-efficiency hydrocyclone at the bottom of the vessel separates the liquid sulfur from the gas. Because of the monolith and hydrocyclone pressure drop, the fuel gas going to the gas turbine is reduced to 350 psia. Sulfur is continuously drained from the sump at the bottom of the hydrocyclone and is stored for shipment.

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[9] Burchell, T. D. et al., "A Novel Process and Material for the Separation of Carbon Dioxide and and Hydrogen Sulfide Gas Mixtures," Carbon Vol. 35(9) pp. 1279-1294 (1997).

Figure 7
Block Flow Diagram – Texaco Radiant Cooler IGCC Plant with Monolith SCOHS Process

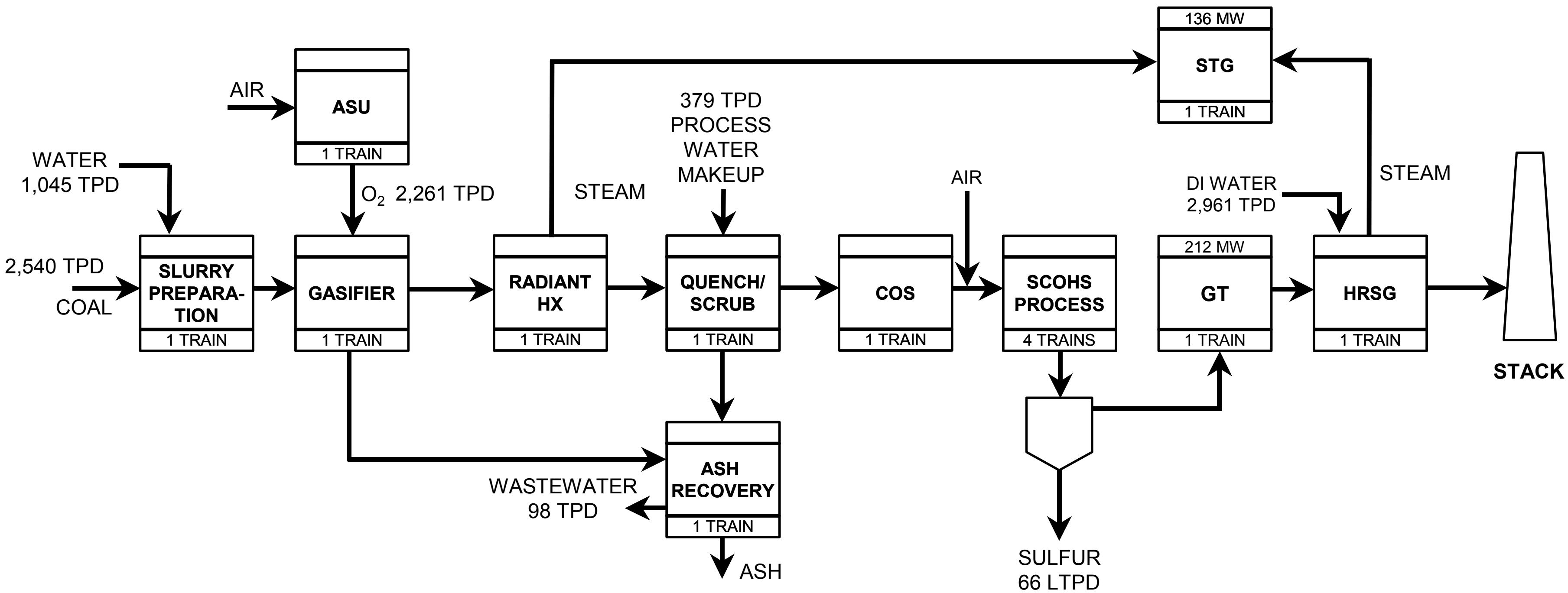

Figure 8
Flow Diagram – Monolith SCOHS Concept

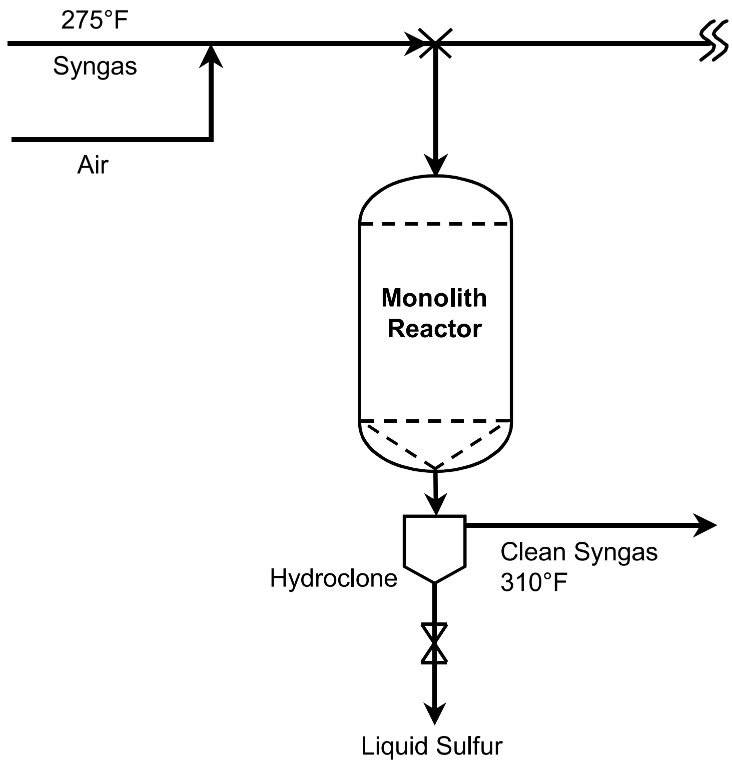

Table 11 summarizes the recommended design parameters for the monolith version of the SCOHS study.

Table 11
Design Basis – Selective Catalytic Oxidation of H_2S for Direct Sulfur Production
The SCOHS Process – Monolith Options

	SCOHS Design Basis Monolith
Syngas	21,933 lb mol/h
Syngas	452,117 lb/h
Molecular weight	20.6
Temperature	275°F
Pressure	402 psia
Sulfur concentration	8,441 ppm
Air stoichiometry	4
Air	265,856 scf/h
Catalyst field packed density	18.7 lb/ft^3
Catalyst loading	N/A
Space velocity	1,000 h^{-1}
Syngas and air	8,139,803 scf/h
Syngas and air	444,659 acf/h
Absorbent catalyst	8,140 scf
Cycle time absorption and regeneration	N/A
Superficial velocity	0.50 ft/sec
Number of vessel pairs or vessels for monolith	2
Vessel inside diameter	12.5 ft
Bed height	33.0 ft
Catalyst life (cycles)	N/A
Catalyst replacement time	N/A
Catalyst replacement rate	N/A
Initial catalyst charge	152 tons

4.1.2 Heat and Material Balance

The monolith SCOHS retrofit of the Tampa Electric IGCC Demonstration Project results in a plant producing a net output of 303 MWe at a net efficiency of 39.2 percent on an HHV basis. Performance is based on the properties of Pittsburgh No. 8 coal. Overall performance for the entire plant is summarized in Table 12, which includes auxiliary power requirements.

Table 12
Tampa Electric IGCC Reference Plant
with Monolith SCOHS Retrofit
Plant Performance Summary – 100 Percent Load

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas Turbine	211,570
Steam Turbine	136,250
Generator Losses	(6,450)
Total	**341,450**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	810
Coal Slurry Pumps	300
Condensate Pumps	140
LP/IP BFW Pumps	30
HP BFW Pumps	2,230
Air Separation Plant	19,230
Oxygen Compressor	9,640
SCOHS Plant Auxiliaries	650
Gas Turbine Auxiliaries	600
Steam Turbine Auxiliaries	350
Circulating Water Pumps	1,670
Cooling Tower Fans	980
Slag Handling	130
Transformer Loss	800
Wastewater Treatment	20
Scrubber Pumps	70
Miscellaneous Balance of Plant	1,000
TOTAL AUXILIARIES, kWe	**38,650**
Net Power, kWe	302,800
Net Plant Efficiency, % HHV	39.20
Net Heat Rate, Btu/kWh (HHV)	8,706
CONDENSER COOLING DUTY, 10^6 Btu/h	634.2
CONSUMABLES	
As-Received Coal Feed, lb/h	211,688
Thermal Input, kWt	772,394
Total Oxygen (95% pure), lb/h	188,436
Water (for slurry), lb/h	87,091

4.1.3 Process Flow Diagram

Figure 9 is the process flow diagram resulting from the monolith SCOHS heat and material balance.

Figure 9 (3 pages) follows.

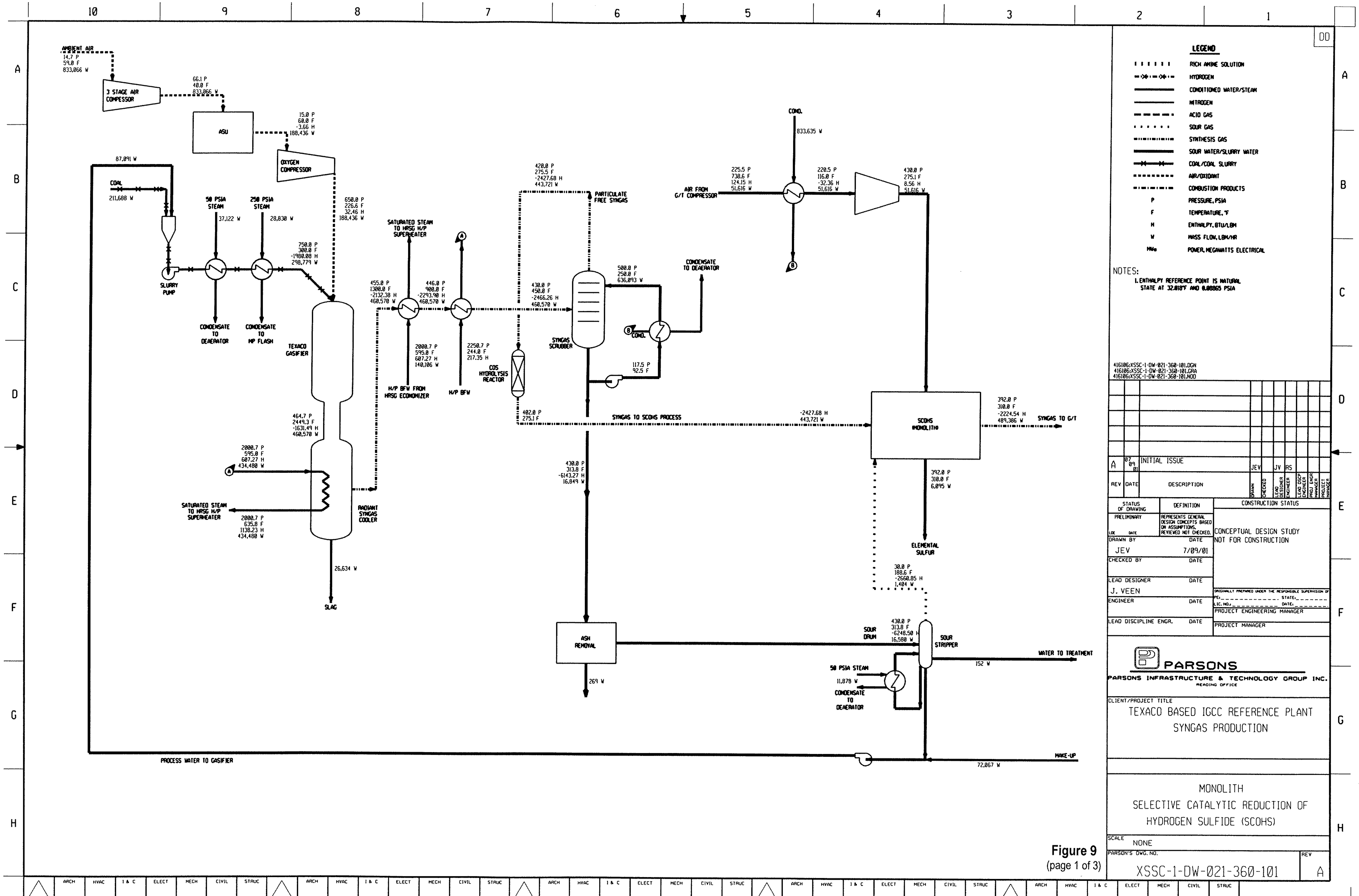

Figure 9
(page 1 of 3)

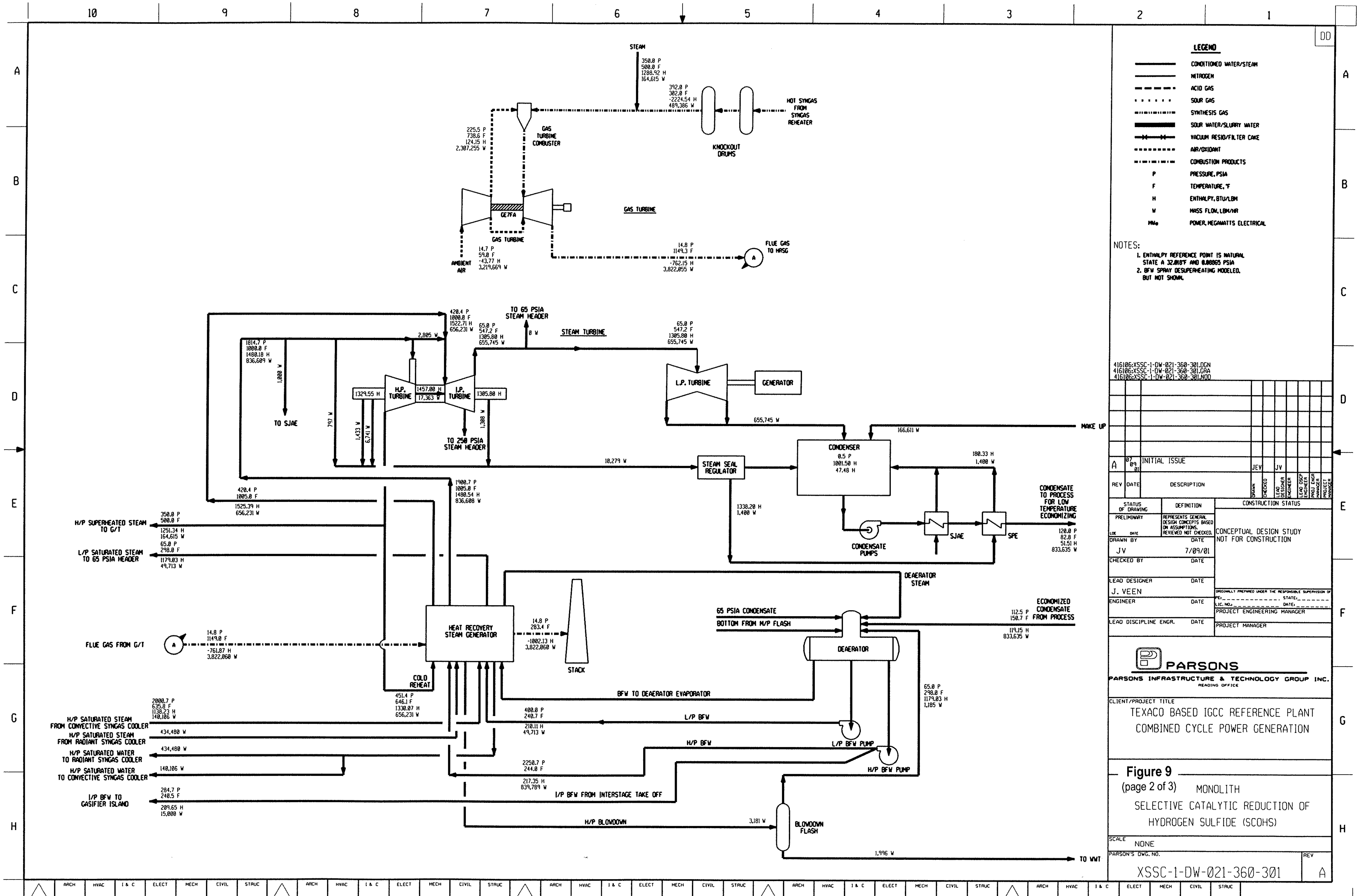

Figure 9 (page 2 of 3) MONOLITH SELECTIVE CATALYTIC REDUCTION OF HYDROGEN SULFIDE (SCOHS)

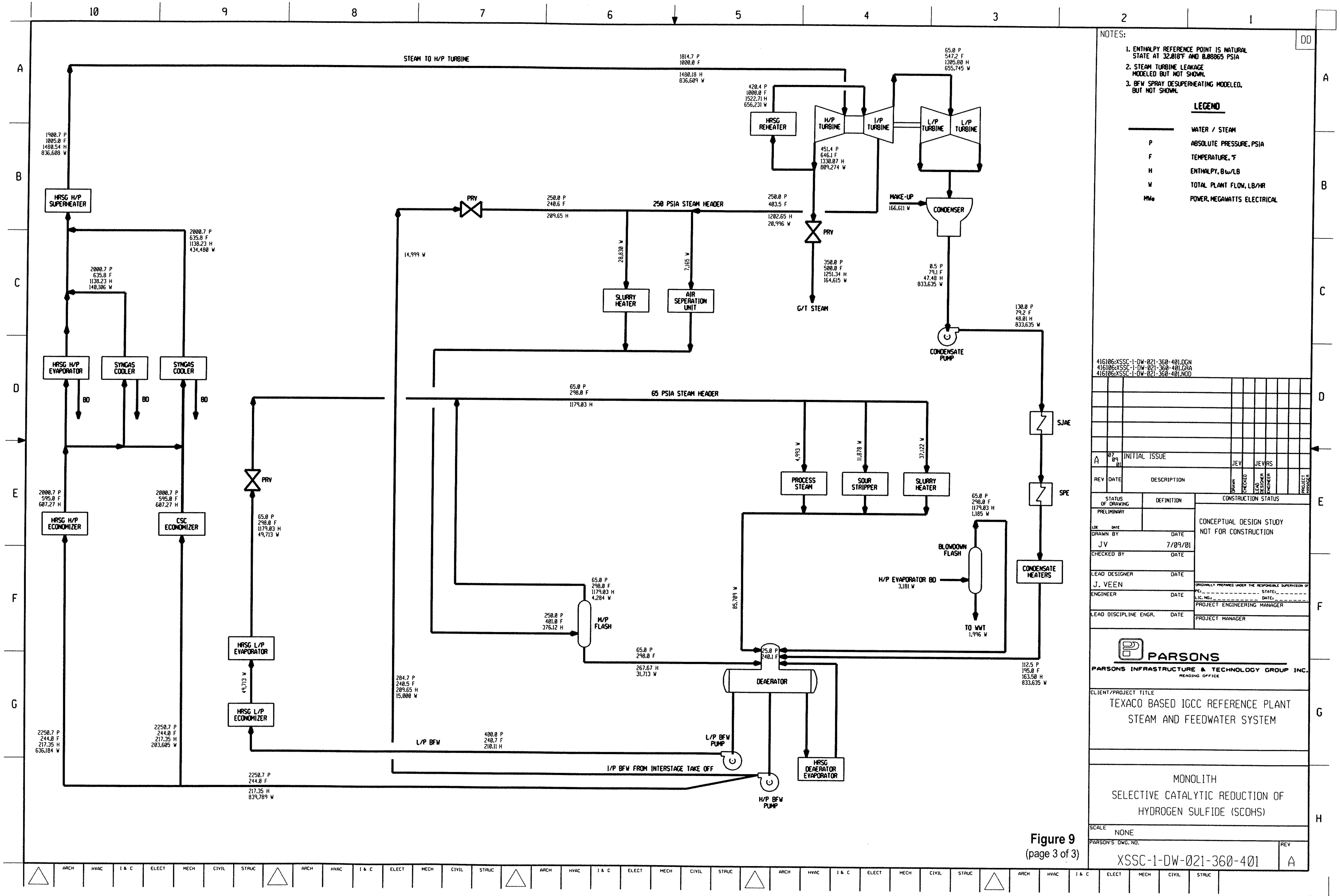

Figure 9
(page 3 of 3)

4.2 COST ANALYSIS

4.2.1 Capital and Operating Cost Estimate

The monolith SCOHS retrofit of the Texaco-based IGCC Reference Plant resulted in a plant that is close in performance and size to the original plant. The overall capital cost difference between the plants is small, notwithstanding the significant cost changes in the sulfur removal and sulfur recovery processes. Table 13 was prepared to show the factored adjustment of capital from the Texaco-based IGCC Reference Plant to the plant with monolith SCOHS retrofit. The initial estimate for the cost of the SCOHS equipment is nominally $5 million. Although the absolute plant total capital requirement changed by less than 4 percent, the monolith SCOHS capital cost per kW is lowered by nearly 9 percent because of higher plant efficiency and higher power production.

Table 13
Capital Cost Changes with Monolith SCOHS Retrofit

		Texaco IGCC Plant Size 287 MWe				Monolith SCOHS Plant Size 303 MWe			
No.	**Account**	**Basis**	**Unit**	**TPC 1,000$**	**$/kW**	**Basis**	**Unit**	**TPC 1,000$**	**$/kW**
1	Coal Handling	209,196	lb/h	$14,931	$52	211,688	lb/h	$15,109	$50
2	Coal Preparation	209,196	lb/h	$14,427	$50	211,688	lb/h	$14,599	$48
3	Feedwater Pumps	2250	kW	$14,883	$52	2,230	kW	$14,751	$49
4	Gasifier	1	Train	$50,251	$175	1	Train	$50,251	$166
4.3	ASU	189,106	lb/h	$39,797	$139	188,436	lb/h	$39,656	$131
5	Gas Cleanup	MDEA/Claus/TGTU		$23,751	$83	Monolith SCOHS		$5,000	$17
6	Combustion Turbine	1	7FA	$54,332	$190	1	7FA	$54,332	$179
7	HRSG	1	Each	$21,878	$76	1	Each	$21,878	$72
8	Steam Turbine	120	MW	$27,660	$96	136	MW	$31,338	$103
9	Cooling Water	518	10^6 Btu	$15,090	$53	634	10^6 Btu	$18,475	$61
	BOP[1] Subtotal	--	--	$65,572	$229	--	--	$65,572	$217
	Total Plant Costs			**$342,572**	**$1,195**			**$330,960**	**$1,093**
	Others			**$37,353**	**$130**			**$36,353**	**$120**
	Total Capital Requirements			**$379,925**	**$1,325**			**$367,313**	**$1,213**

(1) BOP includes ash, accessory electrical, I&C, site, and building systems.

4.2.2 Preliminary Economic Analysis

The cost of electricity (COE) for the monolith SCOHS retrofit was determined by adjusting the COE from the Texaco-based IGCC Reference Plant. Table 14 shows the changes in the components making up the COE. The COE is based on a fuel cost of $1.25 per MMBtu and an annual plant capacity factor of 80 percent. The change in COE with the monolith SCOHS retrofit amounts to a reduction of 8 percent.

Table 14
Change in Cost of Electricity with Monolith SCOHS Retrofit

	Texaco IGCC Plant Size 287 MWe		Monolith SCOHS Plant Size 303 MWe	
	1,000$/y	**$/kW-y**	**1,000$/y**	**$/kW-y**
Capital charge	$52,432	$182.90	$50,689	$167.40
O&M	$12,126	$42.30	$11,626	$38.40
Consumables	$2,332	$8.13	$1,832	$6.05
Sulfur credit @ $47/LT	($898)	($3.13)	($909)	($3.00)
Fuel @ $1.25/ MMBtu	$22,634	$78.95	$22,932	$75.73
Total	**$88,626**	**$309.16**	**$86,171**	**$284.58**
COE @ 80% CF $/MWh	**44.11**		**40.61**	

5. CONCLUSIONS

Conceptual design of an IGCC plant retrofitted with both the fixed-bed and monolith SCOHS concepts indicates that favorable performance and economics can be achieved. The fixed-bed SCOHS results in a COE reduction of 6.4 percent, while the monolith SCOHS results in a COE reduction of 8 percent.

A great deal of developmental work is still needed for the SCOHS process to fully mature. For the simpler fixed-bed process, catalyst development needs to continue to lower COS levels and to allow for an increased temperature of operation. The effect of trace contaminants found in "real" coal-derived synthesis gases and how they affect candidate catalysts is still unknown. Catalysts that possess faster reaction rates are also desirable and would result in smaller fixed-bed reactors. A variety of engineering design parameters still need to be developed to facilitate detailed design and scale-up. These issues cross over into the continuous version of this process as well. The complexity of these systems is sufficient to warrant the construction of a bench-scale unit. Here operational, scale-up, and engineering design parameters would be elucidated on a more acceptable scale. This bench-scale unit would be skid mounted so that it would be possible to operate it off of various coal-derived gas side streams. Based on preliminary economics developed through skid technology, goals could eventually be set that lead to industrial acceptance.